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SYMPOSIUM ON PARTICULATE CONTROL IN ENERGY PROCESSES

Interagency
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Program Report

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16. ABSTRACT The proceedings are a compilation of papers presented at a symposium, cosponsored by EPA's Industrial Environmental Research Laboratory (Research Triangle Park) and the Electric Power Research Institute, and featuring speakers representing the utility industry, research organizations, and government agencies. Purpose of the symposium was to examine the current state of particulate control technology for energy processes, to discuss practical solutions to problems with particulate from Western U.S. coals, and to consider recent progress in high-temperature/high-pressure energy processes. Control of particulate emissions from conventional power generation is well developed and reasonably effective. However, some recent technological developments--increasing use of low-sulfur, high-ash Western U.S. coals, and new high-temperature/high-pressure processes--have increased the difficulty of achieving effective particulate control.

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INTRODUCTION

Methods of control of particulate emission from conventional electric power generation facilities are well developed and reasonably effective. However, several recent developments in combustion technology have significantly increased the difficulty of particulate collection. There is an immediate problem with the rapidly increasing use of low-sulfur, high-ash Western coals. Because of the high mineral content of these coals, ash is produced at higher rates than for high-sulfur (Eastern) anthracite or bituminous coals. Compounding the problem, the low sulfur content decreases the effectiveness of electrostatic precipitators. The most severe problems arise when attempting to switch existing power plants (with precipitators sized for high-sulfur coal) to low-sulfur coals.

A longer-range problem with particulate collection exists with several high temperature and pressure energy processes now being developed. In order to obtain maximum efficiency of power generation from coal gasifier or fluidized bed combustion facilities, it is necessary to remove particulates from hot, high pressure streams upstream of the turbines.

The "Particulate Control in Energy Processes" symposium, held in San Francisco on May 11-13, 1976, devoted six sessions, two panel discussions, and some 22 papers to consideration of these problems and their solutions. This volume is a collection of the papers presented at the symposium.

The symposium was sponsored by the Particulate Technology Branch of the U.S. Environmental Protection Agency (Research Triangle Park, North Carolina), and by the Electric Power Research Institute (Palo Alto, California).

The symposium was hosted and the proceedings produced by the Aerotherm Division of Acurex Corporation (Mountain View, California).

I wish to express my appreciation for the help and support of the sponsoring organizations, and most particularly to Dr. Dennis C. Drehmel of the EPA and Mr. Walter Piulle of EPRI for the many hours they spent working with me to assure the symposium's success.

David E. Blake
Symposium General Chairman
Aerotherm Division
Acurex Corporation

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LOW SULFUR COAL AND THE PARTICULATES PROBLEM

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LOW SULFUR COAL AND THE PARTICULATES PROBLEM

All of the participants in this symposium are involved, in some manner, in a classic environmental problem. In essence, the problem we are addressing is that, in controlling one form of pollution (in this case sulfur oxides emissions), we have found increased difficulties in controlling another form of pollution (particulates). This is a classic environmental problem in that it brings to mind all of the buzz-words which have embellished environmental debate in the past. The problem involves trade-off analysis. It is based on health effects studies and on emission standards. And, above all, it requires a comprehensive, holistic and integrated solution with appropriate control technology development. As I said, it's a classic environmental problem.

DDT BROUGHT THE HOUSE DOWN

In its complexity, this problem reminds me somewhat of another classic environmental problem which came to light several years ago on the other side of the globe. The story, which is said to have taken place in Borneo, involved a government effort to rid isolated native tribes and their thatched houses of lice and malaria. To do so, the government sent teams to the villages to spray the huts with DDT. The effort was a stunning success. For the first time in history, the natives stopped itching and malaria was curbed.

In a short time, however, an unexpected problem developed. It seems that the lizards that shared the thatched huts with the natives were eating the poisoned insects and concentrating the DDT. And the village cats were eating the sick lizards and then expiring in droves. It didn't take long before the jungle rats, realizing their opportunity, invaded the villages and terrorized the natives. In response, the government's Royal Air Force airlifted hundreds of cats into the villages to drive back the rats. Unfortunately, that wasn't the last of the problems. It seems that, without their natural lizard predators, the worms that dwelt in the thatched roofs of the huts proliferated and literally ate the natives out of house and home. Entire villages collapsed. Thus the natives were left without homes to call their own and surrounded by a bunch of strange cats. I'm sure you all know the feeling.

The moral of my story is that, when environmental problems, such as we are addressing in these proceedings, arise, our solutions must be as far-sighted as possible. Rather than just resolving today's crisis, we should try as hard as possible to develop the kind of solutions and control technologies which will also hold the greatest potential for solving tomorrow's problems as well.

The particular problem that this symposium is addressing is the need for ways in which to utilize low-sulfur coal in an environmentally compatible manner. Some of the more far-sighted discussions of solutions to this problem are presented in the papers on high-temperature, high-pressure particulate control technology -- a capability which is essential to the successful development of promising second-generation energy technologies.

EPA'S PARTICULATE CONTROL R&D EFFORTS

The remainder of this paper will be devoted to placing in perspective EPA's research and development role in particulate control, and to presenting some of our significant problems and identifying the major programs which are attempting to find solutions to these problems.

As the nation strives to reduce dependence on foreign energy sources, increased consideration is being given to the energy-related environmental problems arising from domestic energy resource development. Accelerated efforts to expand existing energy supplies and to develop new energy sources are accompanied by accelerated efforts to anticipate, plan for, and control the more serious adverse health and ecological effects resulting from such energy development.

Since its establishment in 1970, the Environmental Protection Agency has been heavily involved in energy-related environmental research efforts, including the development of pollution control technology, supportive of the EPA's statutory responsibilities. Since Fiscal Year 1975, the EPA has also had the added responsibility of planning and coordinating the entire 17-Agency Federal Energy/Environment Research and Development Program. Through the Interagency Program to date, the EPA has planned and implemented more than a quarter of a billion dollars of energy-related environmental research and control technology development. Responsible for this centralized coordination of the Interagency Program is the Office of Energy, Minerals and Industry, within EPA's Office of Research and Development. This organization pursues two basic goals -- to provide a focus for EPA's own Environment/Energy and Industry-related R&D efforts, and to serve as the coordinator of the entire Federal Interagency Program.

EPA'S GOAL -- CLEAN USE OF COAL

EPA's mission in this area recognizes that timing is critical in the overall relationship between energy development and environmental protection. Efforts to expand use of domestic supplies will focus on the nation's two primary fuel resources -- coal and nuclear energy. Through the 1970's and early 1980's, coal use in particular will be increased. This will occur chiefly through direct combustion in power plant and industrial boilers, but only if technologies to control emissions of particulates and other hazardous pollutants are developed and applied. Thus, Environment/Energy R&D on near-term coal use centers on our continuing program to advance environmentally acceptable ways to extract and utilize coal. To the degree that these control technologies can be implemented rapidly, there will be early environmental, economic, and social benefits from such R&D efforts. For this reason, funds have been weighted heavily in the FY 1975 and FY 1976 budgets towards facilitating near-term coal use.

The Clean Air Act is the basis for our research into and development of flue gas cleaning technologies. This R&D is focused on identifying a series of standards which can be implemented within the constraints of sound engineering practices and economic viability.

We are aggressively pursuing research on particulate control because of the potential widespread impacts of accelerated coal resource development and usage. Particulate control needs and required technologies are, to a large extent, dependent upon the type of coal to be used, the combustion process involved and the location of such usage. The bulk of coal usage is in the Eastern States, but the bulk of low sulfur coal is in the Western States. See Charts One and Two for the distribution of domestic reserves now available. There is almost equal distribution of coal between Eastern and Western states, but there is five times as much low-sulfur coal in the West. There is also a greater proportion of strip-mineable coal in the West. For Eastern users the coal may be obtained by deep mining low-sulfur Eastern coals, by desulfurizing strip-mineable, high-sulfur coal, or by shipping Western coals to the East. These coals, whether of high or low sulfur content, will be available for utility and industrial use for many decades.

THE SULFUR/PARTICULATE TRADEOFF

Typical of the problems facing the EPA and industry is the effect on electrostatic precipitator operations of conversion to low-sulfur coal usage. The electrostatic precipitator (ESP) is the most common device for controlling the particulate emissions from large coal-fired utility boilers. Those familiar with the power industry know that in an electrostatic precipitator, dust suspended in the gas stream is electrically charged and passed through an electric field which causes the particles to migrate toward a collection electrode.

The particle collection efficiency of an ESP is, in part, a function of the resistivity and particle size distribution of dust being collected.

The fact that fly ash resistivity, and thus precipitator performance, are related to coal sulfur content has long been recognized. Correlations have been developed between ESP performance and coal sulfur content. It is clear that a given electrostatic precipitator will have higher collection efficiencies when collecting fly ash from the combustion of high-sulfur coal than when collecting fly ash from the combustion of low-sulfur coal.

CHARTS ONE & TWO: COAL RESERVES IN THE UNITED STATES

		DEEP-MINABLE	*STRIP-MINABLE*	*TOTALS*
CHART 1. EASTERN STATES	RESERVES			
	%	83	17	100
	BILLION-TONS	169	33	202
	RESERVES			
	≤ 1% SULFUR	33 BILLION TONS		16
	1.1 TO 3.0%	33 " "		27
	> 3%	55 " "		40
	UNKNOWN	82 " "		16
CHART 2. WESTERN STATES	RESERVES			
	%	56	44	100
	BILLION-TONS	131	103	234
	RESERVES			
	≤ 1% SULFUR	167 BILLION TONS		71
	1.1 TO 3.0%	38 " "		16.0
	> 3%	11 " "		5
	UNKNOWN	18 " "		8

The switch to a low-sulfur coal, as necessitated by efforts to achieve or maintain sulfur oxides emissions standards, can increase particulate emissions by more than a factor of 10. As can be seen from Chart Three, such an increase in particulate emissions on a large scale basis would result in unacceptable air quality impacts. Major programs to identify and evaluate alternatives to overcome this problem, or to establish the viability of other control strategies, are being implemented by our research program (see Chart Four).

In parallel with our efforts to develop the necessary control technologies, an Interagency Program effort is being pursued to identify health effects, ecological fate and effects, and atmospheric effects of particulates. In addition to the EPA, major Interagency Program efforts in this area involve NIOSH, ERDA and NIEHS. Through our role in coordinating such interagency efforts, we are becoming increasingly aware of the need for continuing research into the effects of particulates as well as into the development of control technology options.

FINE PARTICULATES -- A PERVASIVE PROBLEM

Specific program focus is on the range of smaller particulates (see Chart Five). Fine particulates appear to pose a far greater health hazard than do the more easily controlled large particulates because, in contrast to coarse particles, they can bypass the body's respiratory filters and penetrate deeply into the lungs. Fine particles released into the atmosphere remain airborne for extended periods of time, obstruct light, cause the limited visibility typical of air pollution, haze and smog and are added to the planetary particulate loading which may, according to recent studies, have a significant climatic cooling effect.

Such fine particles have also been identified as transport vehicles for gaseous pollutants. The health hazards of fine particulates are intensified by the tendency of emissions from pyrometallurgical and combustion processes to condense as chemically active fine particles. Many toxic and potentially hazardous compounds are also emitted as fine particulates. Such particulate matter, formed in the

CHART THREE
PARTICULATE EMISSION ESTIMATES
EMISSIONS (1000 TONS PER YEAR)

	1970		1975	
SOURCE CATEGORY	POTENTIAL	ACTUAL	POTENTIAL	FULL COMPLIANCE
○ COAL-FIRED STEAM ELECTRIC POWER PLANTS	34533	4188	47530	914
○ INTEGRATED IRON AND STEEL MILLS AND COKE PLANTS	15259	2800	18532	521
○ PORTLAND CEMENT PLANTS	10643	906	12291	69
○ OTHER	29017	19606	34824	10896
○ ALL SOURCES	89452	27500	113177	12400

CHART FOUR
FINE PARTICULATE PROGRAM

- **MEASUREMENT TECHNIQUE DEVELOPMENT**
  - GOAL IS TO PRODUCE DEVICE WHICH IS CONTINUOUS AND ACCURATELY MEASURES FRACTIONAL EFFICIENCY
- **ESP PROGRAM**
  - CHARACTERIZATION FOR SEVEN SOURCES
  - MATH MODEL IS DEVELOPED
  - CAPABLE OF HIGH EFFICIENCY. IF NO RESISTIVITY PROBLEM
  - CONDITIONING BEING EVALUATED
- **SCRUBBERS**
  - EIGHT DEVICES TESTED ON A VARIETY OF SOURCES
  - PERFORMANCE DEPENDENT ON PRESSURE DROP
  - TCA SCRUBBER UNCHARACTERISTICALLY EFFICIENT
  - FORCE/CONDENSATION SCRUBBERS UNDER DEVELOPMENT
- **FABRIC FILTERS**
  - TESTING FOR THREE SOURCES COMPLETED
  - EFFICIENT DOWN TO 0.3μ
  - PROGRAM AIMED AT INCREASING APPLICABILITY AND OPERABILITY
- **NEW IDEAS/NOVEL DEVICES**
  - 40 NEW CONCEPTS EVALUATED; 9 APPEAR PROMISING
  - 30 NEW DEVICES EVALUATED; 10 APPEAR PROMISING
- **HIGH TEMP./HIGH PRESSURE CT**
  - FBC AND LOW-BTU GASIFICATION NEED CT
  - PROGRAM AIMED AT DEFINING REQUIREMENTS AND EVALUATING CONCEPTS AND DEVICES

CHART FIVE
FINE PARTICULATE CONTROL
PROGRAM-BACKGROUND

–FINE PARTICULATES ARE HEALTH HAZARDS BECAUSE:

- ○ AIRBORNE FOR EXTENDED TIME PERIODS
- ○ PENETRATE DEEPLY IN LUNG
- ○ ACT AS TRANSPORT AGENTS FOR OTHER POLLUTANTS

–EPA R,D & D PROGRAM INCLUDES

- ○ MEASUREMENT TECHNIQUE DEVELOPMENT
- ○ IMPROVEMENT AND CHARACTERIZATION OF PRESENT TECHNOLOGY
    - ESPs
    - SCRUBBERS
    - FABRIC FILTERS
- ○ NEW IDEAS/NOVEL DEVICES
- ○ HIGH TEMP./ HIGH PRESSURE CT

atmosphere from chemical reaction and condensation, is called secondary particulates. Such secondary particulates raise added problems and issues. The phenomena associated with the formation and transport of secondary particulates makes it difficult to relate atmospheric particulate pollution levels to specific sources. This difficulty has hampered the development of effective fine particulate control strategies and the establishment of meaningful fine particulate emission standards. The control of these secondary forms of particulates must be through control of their precursors, and it is thought that primary emitted particulates play an important role in the formation cycle.

Hence, the information which is being developed through our health and ecological effects research efforts continues to add support to the need for a control technology capable of controlling fine particulates while, at the same time, allowing for effective control of sulfur emissions as well. The more we learn about the pathways and hazards of particulate pollution, the more our need for an effective set of control technologies is manifest. Like the Borneo natives, we require control technologies which are relevant to, and compatible with, the entire range of environmental conditions that we have identified.

CONCLUSION

In conclusion, there are several points which I would like to emphasize:

First, in the near term there will be no energy panaceas. Our new sources of energy will look a lot like our old sources of energy and will employ technologies with which we are already familiar.

Second, the protection of public health and welfare requires that the emission of both sulfur oxides and of particulate matter be controlled at the source.

And finally, in those instances where low-sulfur coal is to be used as an energy source in order to control sulfur emissions, there is a technological conflict between the requirements imposed by sulfur oxide standards and the ability of the traditional particulate control devices to meet particulate control needs.

With these points in mind it is obvious that particulate control in energy processes is a topic deserving of immediate and intensive study. Solutions to the existing problems, especially in regard to low-sulfur coal usage, require concentrated effort not only on the part of governmental agencies but also in cooperation with utilities and other consumers of the fuel. It is fortunate that the natural and understandable reluctance of these industries to give rapid trial and endorsement to changing technology has begun to abate. Such tempering of attitude is beneficial and will assuredly lead to quicker and probably better solutions to the present problems. We have a common problem. We should work together to develop a common solution.

LOCATION OF LOW SULFUR COALS AND WESTERN POWER GENERATION NEEDS

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INTRODUCTION

Assessments of the coal supply in the ground have been published by the dozens in the last years. These assessments come in all shapes and varieties, some reporting dire shortages within a few years to no likely shortage within a few million years. When one analyzes these forecasts, one can come to the conclusion that each may be correct. The variations depend entirely on the different assessments of the coal availability measured with geologic, technological, economic and political parameters. Within the last few years, since the enactment of the Clean Air Act, a new parameter has been added, the quantitative amount of qualitative sulfur content of the unmined or in-situ coal. The first set of parameters are difficult enough to comprehend and make judgment upon to arrive at a conclusion. The last parameter concerning the sulfur content of coal upset the whole trend of reasoning.

To most of us, a resource is a thing or action we can turn to in time of need or emergency. To a geologist the word indicates a concentration of raw inplace coal in such form that economic extraction is currently or potentially feasible. But the common man's concept of resource, when it comes to coal, includes only "reserves." Reserves are assumed to be ready for extraction and consumption, and can be drawn upon to take care of a need or an emergency. Undiscovered seams of coal included in resource are not available for use, they are only presumed to be present in the earth's surface.

Coal Resources and Reserves: Quantity

The most serious problem in determining both coal resources and recoverable reserves is that much of the basic technical field work in the public domain was performed many years ago. Many estimates are founded on work done in the early part of this century without the advantage of modern techniques and instrumentation and thus are often lacking the degree of refinement and specificity necessary in assessment of potentials for the large scale, costly developments of the present. An intensive effort in this regard is to be carried out by the present programs of the U. S. Geological Survey and the U. S. Bureau of Mines. The Geological Survey is primarily responsible for determination of the nation's resources, while the Bureau of Mines provides estimates of recoverable reserves. The data on coal resources and reserves resulting from the work of the Geological Survey and the Bureau of Mines represent the only authoritative accounting of the nation's coal deposits, and these data will be used to describe the situation apparent at present.

The United States Geological Survey estimates the total remaining coal resource in the United States with certain specifications, namely beds of bituminous coal and anthracite 14 inches or more thick and beds of subbituminous coal and lignite 2-1/2 feet or more thick with over burden depths of 3,000 and 6,000 feet.

The resulting estimate of the amount of coal resources -- $3,968 \times 10^9$ (billion) tons -- is a theoretical number, some of this coal is mineable under economic conditions, but most of the coal cannot be extracted with present mining techniques.

The total remaining coal resources of the United States to a depth of 3,000 feet as determined by mapping and exploration is about 1,580 billion tons.[1] The total comprises, in rough terms, the following:

- 686 billion tons bituminous coal
- 424 billion tons subbituminous coal
- 450 billion tons lignite

Clearly this is a substantial amount. However, as will be shown in this section, not all the coal resource can be recovered, thereby reducing the quantity of usable reserves. Additionally, as shown in the next section, the different heating values and quality of coals further reduce the magnitude of the available coal reserves.

In order to be considered targets for development, coal seams must be of relatively high quality (ideally, at a minimum, high heating value and low sulfur content) occurring in relatively thick, uniform, and extensive beds that are located near enough to the earth's surface to permit application of economic mining technology. Thickness of coal seams is a most important parameter, because it controls the lateral extent of operations and strongly influences operating costs. Figure 1 shows the relationship of coal seams thickness to area required to account for 1 million tons of reserves. It may be seen that a 4 foot coal seam (common in certain eastern coal fields) requires 140 acres for 1 million tons, whereas a 12 foot seam (common in certain western coal fields) requires only 50 acres. Clearly, other things such as coal character and quality being equal, it would be preferable to work in the more concentrated deposit. This is not always possible, however, because of consumer requirements for coal properties and quality.

As a part of planning for Project Independence, a new evaluation of the nation's coal deposits were made by the U.S. Bureau of Mines and Geological Survey. A new term was coined; the "reserve base", defined as "the quantity of in-place coals calculated under specified depth and thickness criteria." This new term is intended to be intermediate between resources (the total stock of coal in the ground) and reserves (the quantity of coal actually recovered).

The coal reserve base uses more realistic criteria for estimating. It includes only coal reserves in the measured and indicated resource category; in beds 28 inches or more thick for bituminous and anthracite and 60 inches or more for subbituminous and lignite; in the 0 to 120 feet overburden category for lignite, which is considered suitable only for strip mining, and in the 0 to 1,000 foot overburden category, which are deemed to be suitable for strip, auger and underground methods.

The "demonstrated coal reserve base", includes the measured and indicated reserve categories, as in the coal reserve base, but based on a high degree of identification and engineering evaluation with emphasis

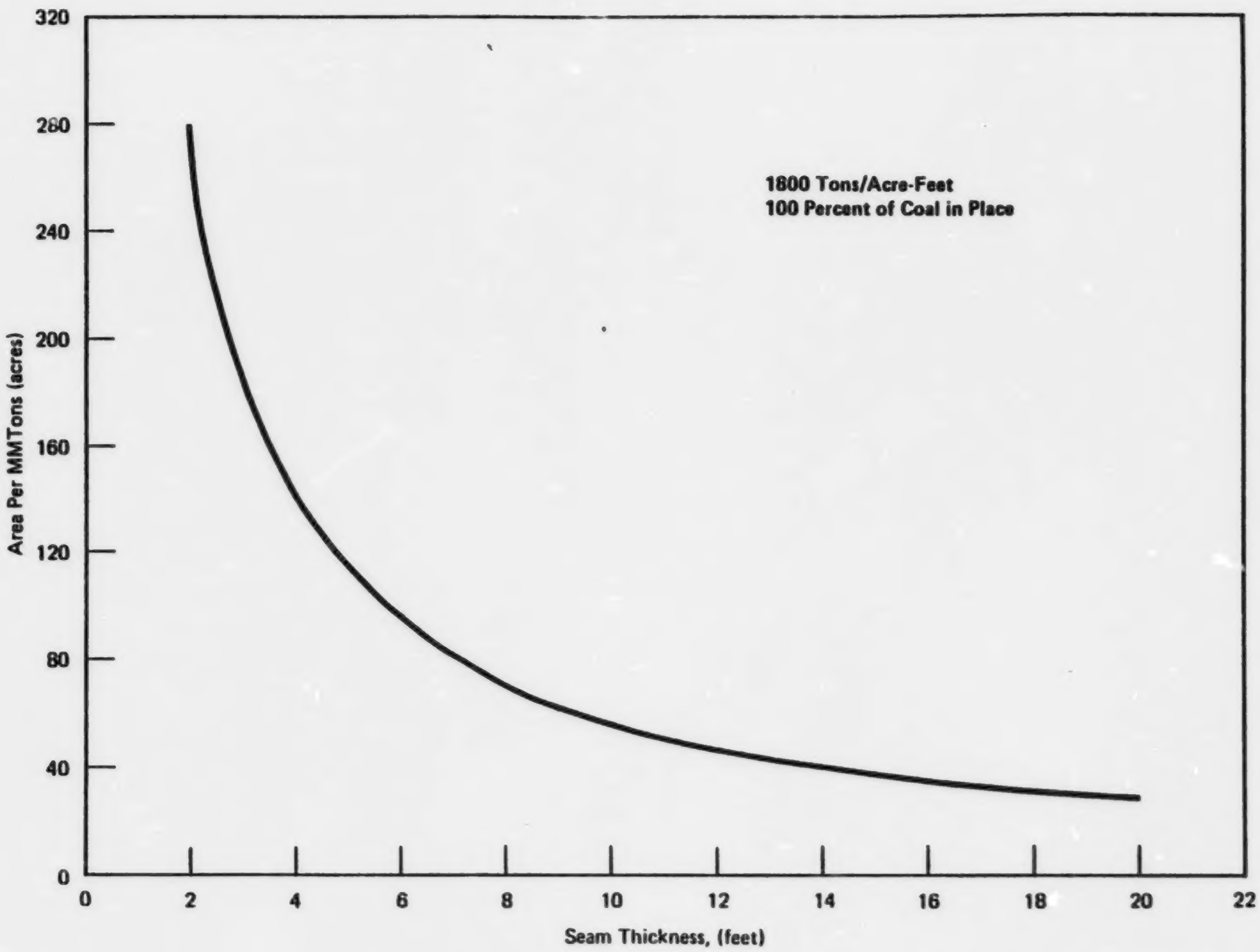

Figure 1 Relationship of Seam Thickness to Area per Million Tons

given to a high degree of certainty that the coal is technically and economically mineable at some time. No consideration was given the marketability or competitiveness of one seam of coal against another. In effect it is assumed that all of the coal can eventually be mined if the mining technology is perfected to extract it and there is a customer demand great enough that it is willing to pay the price.

The "reserve base" data represent the most recent estimate of the quantity of U.S. coal deposits which may experience development in the future. Table 1 presents "reserve base" data together with the authors' estimate of recoverable reserves for both underground and surface mining.

Tables 2 through 5 show, for each of the four principal coal ranks, the following information by state:

(a) total estimated identified resources
(b) indicated reserve base, by method of mining
(c) recoverable reserves, by method of mining, and total
(d) recoverable reserves as percent of identified resources

The tables are, with one exception, compiled from available information. The exception is that an estimate of recoverable reserves by underground mining was prepared for each rank of coal. Note that the reserve base represents 100% of the coal in-place; no allowance for incomplete recovery has been made. Actually, the amount of coal that can be recovered from a given deposit varies from about 25% to 90% of the coal in-place. Therefore, in order to arrive at a reasonable estimate of the amount of recoverable coal, the "reserve base" data must be reduced by an appropriate amount to allow for incomplete recovery.

All of the schedules of coal resources and reserves presented so far are estimates of the coal in the ground and not necessarily mineable under unspecified circumstances.

In general terms, the Bureau of Mines uses a round recoverability factor of 50%. The recoverability factor is defined as the percentage in the reserve base that can be recovered by established mining practices. The 50% recovery factor was established from past mining records for underground mines in the eastern states. In the East, most of the mining to date has been accomplished in areas where multiple seams of coal are present. For economical reasons, it can be conceded that the prime seam of coal was or is being mined first. Ideally, this procedure leaves the seams above and below undisturbed. The Bureau of Mines counts undisturbed seams as mineable, which may not be necessarily true. The seams above a mined out seam, if the interval between the two is not sufficiently thick, may be fractured and subsided to such an extent that the seam is not mineable under any conditions. A similar situation may develop with coal seams below the mined out seam. Water may seep through fractures in the interval cover material to such an extent that the roof is not supportable, again the seam is lost for mining, but still considered as a coal reserve.

Seams of coal under populated areas, Federal and state owned forests, parks, reservations, navigable rivers and streams, etc., are not legally mineable areas, but these areas are considered in the coal reserve count. Neither is the land around oil and gas wells mineable, large blocks of coal have to be left standing to prevent the hazard of oil and gas seepage.

The various geological surveys that have worked many years in compiling geological data because of limited funds they have had to use limited data, coal seam outcrop information, water well drilling records and any other available data to make their judgement on the continuity of a seam of coal. Usually, this type of available data is very sparse, and has to be extrapolated for very large areas. Such data, many times, does not show the thinning out of a seam of coal, seam splitting, geological faults and voids and many other conditions, but the factors are not considered when appraising an area.

Very large tracts of land were assembled in the past for a mining operation which may have been successful in mining all of the available recoverable prime seam coal and leaving the remaining seams in place. In many cases, these large tracts were later subdivided and sold off as farms, building lots, etc., making the reassembling of these lands into a mineable tract impossible; again this coal is lost as a mineable reserve. In the West, where very thick seams of coal are found, Federal and state laws restricted the mining of coal which may disturb dry river beds, alluvial plains, ceremonial grounds, etc., which eliminates many more tons of coal reserves. The following analysis represents an initial attempt to take these factors into account in arriving at an estimated actual recoverable coal reserves.

Recoverable reserves for underground mining were estimated by applying a recovery factor to the reserve base. Although the recovery of coal reserves in a given deep mine may be about 50%, there are substantial areas between mines, under cities, in pockets too small for economic recovery, too badly faulted, etc., where mining is either impossible or otherwise precluded. The effect of these and other conditions is to lower the recoverable reserves. To allow for the influence of the above factors, it is assumed in this analysis that coal recovery by underground mining from the "reserve base" will be, at a maximum, only about 30% of the total coal in place. Clearly, this recoverability factor is uncertain; it could go as low as about 20% or as high as about 35% of the total coal in place. It is considered unlikely, however, as suggested by the U.S. Bureau of Mines, that recoverability from the "reserve base" would be as high as 50%. Table 1 shows that the total estimated recoverable coal through underground mining is nearly 85 billion tons, with about two-thirds of the total comprised by eastern and midwestern bituminous coal, about one-third subbituminous coal, and minor amounts of anthracite. This estimate, recognized to be conservative, could be increased slightly by applying a different recoverability factor. In view of the several uncertainties regarding estimating coal reserves, it is believed prudent to be conservative at the present so that developments of finite resources of coal can be carefully planned and evaluated in advance.

For surface minable reserves, the estimates of the U.S. Bureau of Mines[2] have been adopted (Table 1). These data show that the estimated recoverable coal through surface mining is roughly 45 billion tons. More than half of the total is subbituminous coal, with nearly one-third bituminous coal and the remainder lignite.

The recovery of coal from reserves through surface mining may also be less than the commonly quoted estimates of 85% or 90% or more.[3] While these magnitudes may apply to a given mine, they may not apply to an entire coal field. Data in Table 2 through 5 permit an assessment of the apparent resource recovery percentage in surface mining. Assuming that the strippable reserve base includes all coals that may be extracted through surface mining, and assuming further that the separately estimated recoverable strippable reserves represent the portion of the reserve base that can be actually produced, then it is possible to calculate an indicated recovery percentage for surface mining for each rank of coal, by state.[4] Bearing in mind the shortcomings of the approach used, the data have been employed to calculate an "indicated recovery percentage" for strippable reserves of different rank coals, by state (Table 6). The table shows that there are few states where the indicated recovery percentage is even close to the commonly stated "norm" of 85% to 90% or greater recovery. In most major coal-producing states, only about one-quarter of the reserve base is estimated to be recoverable (West Virginia, Illinois, Kentucky). Stated another way, if the reserve base figure had been employed together with the "normal" 85% to 90% recovery factor to derive recoverable strippable reserves, values would have resulted which would be much larger than those determined by Bureau of Mines measurements. The actual recoverable strippable reserves, which largely are only about 45% of the reserve base, and will clearly require careful management and conservation if they are to satisfy projected demands.

The foregoing suggests that the coal reserves recoverable from the reserve base are:

Mining Method	Commonly Estimated Recoverable Reserves, Percent of Reserve Base	Indicated Recovery Percentage, Percent of Reserve Base
Underground	50	30
Surface	90	45

In addition, it will be essential to extend the reserve base by more complete geological knowledge and through the perfection of new and improved technology.

The total estimated recoverable coal resources as indicated by the above data is 130 billion tons, or about 8% of the total U.S. coal resources. Clearly, the quantity of recoverable reserves is still substantial. However, as will be discussed next, the coal quality factors limit the potential use of recoverable coal reserves.

Coal Resources and Reserves: Quality

To an electric utility coal consumer, the heating value of coal (expressed as the number of Btus per pound or per ton) is the important parameter, not merely tonnage alone. In producing a given amount of heat for production of electricity, a utility would have to use significantly larger amounts of low Btu coal than of high Btu coal. Unfortunately, in consuming the additional tonnage of low Btu coal to make up the Btu difference, the sulfur content of the additional tonnage is also emitted, and consequently the effective sulfur content of the coals is increased. Employing a standardized heating value for coals used by electric utilities, Rieber[5] found that "conventional estimates of both known resources and recoverable reserves of low sulfur coal are grossly overstated." In particular, Rieber's analysis shows a small increase in the estimates of bituminous coal resources/reserves, and to a large reduction in the estimates of subbituminous coal and lignite. A significant portion of U.S. coal resources and reserves which are normally considered to be low in sulfur content are reclassified to higher sulfur categories. For example, known recoverable reserves in the lowest sulfur category ($\leq$ 0.7 wt. percent sulfur) are reduced from a conventional estimate of 68.2 billion tons to 16.4 billion tons on a consistent Btu-sulfur adjusted basis (Table 7). Should coal demand be such as to require a doubling over present levels by 1985, the cumulative production would be on the order of 20 billion tons. With conventional reserve estimates there would be close to 50 billion tons of coal remaining, even after such expanded production. However, under the revised, standardized reserve estimates, known recoverable reserves of lowest sulfur coal would be completely exhausted before 1985, and there would even be a deficit!

The reduction in conventional reserve estimates is concentrated in the western states (Table 7), where nearly 85% of these reserves are shifted to higher sulfur categories through the standarization procedures. The reserves most affected are the lower heating value coals (subbituminous and lignite) which are present in the largest tonnages.

These data are of exceptional significance for long-range policy planning. There seems to be no practical alternative to use of bituminous coals of high heating value and high sulfur content, simply because of the scarcity of low sulfur coal of equivalent heating value. Revision of air quality standards would be the most immediate way to achieve continued use of certain high sulfur coals and expanded use of others, with or without flue gas desulfurization devices. Alternatively, high sulfur coals could be gasified or liquefied to produce "clean fuels;" it remains to be seen whether such substitute fuels would be restricted to markets other than electric utilities because of the inherent properties of the converted fuels and shortages of natural fluid hydrocarbons.

Table 8 through 11 presents data on recoverable reserves of the several coal ranks by state. The average heating value and sulfur contents as reported by the U.S. Bureau of Mines were used to calculate the sulfur contents per million Btus. The sulfur/heating value data were compared with recoverable reserve data to arrive at an assessment of potential recoverable reserves. Only those states having relatively large recoverable reserves of low sulfur coal per unit of heating value were judged to have significant development potential. These estimates are upper limits, with the actual production realized being determined by individual coal seams and probably less than these amounts.

Table 12 summarizes the potential recoverable reserves for each rank by sulfur content per million Btu. Roughly half of the total recoverable reserves are bituminous coal, and roughly two-thirds of this has more than 1.5 lbs S/MMBtu. The next largest amount is subbituminous coal, all of which is less than 1.5 lbs S/MMBtu. The relatively smaller recoverable reserves of anthracite and lignite are all less than 1.5 lbs S/MMBtu.

While about one-third of the total potential recoverable reserves have less than 0.7 lbs S/MMBtu, nearly three-quarters of this lowest sulfur coal occurs in the western states. The relatively remote location of these reserves and the uncertainty about their development resulting from pending litigation could delay the pace of their development. About one-quarter of the total potential recoverable reserves are in the next lower sulfur category; again roughly three-quarters of these reserves occur in the West and their early development is not at all assured. The remaining reserves, while occurring in the East, are greater than 1.5 lbs S/MMBtu. Although many of these reserves have been developed, it is not at all apparent that the coal could be used without treatment of some sort in electric utility plants under present environmental control requirements.

Tables 13 and 14 present a comparison of assessed recoverable reserves against projected production. The data show an optimistic estimate of future coal production to 1985 (assuming that new production is in addition to present production which is depleted by about one-quarter because of mine retirement or exhaustion). Only about one-quarter of the 1985 production would require at least some treatment prior to use. The remaining half of total 1985 production is quite high in sulfur content and would require rather extensive treatment in order to permit use of these coals in power generation.

The estimated 1985 production of bituminous and subbituminous coal is distributed among broad geographic regions as follows:

Estimated 1985 Production (million tons)
(lbs S/MMBtu)

Region	<0.7	0.7 - 1.5	> 1.5	Total
East	39.2	135.9	487.8	662.9
West	209.8	94.6	9.5	313.9
Total	249.0	230.5	497.3	976.8

The data indicate that two-thirds of total 1985 production is expected to come from east of the Mississippi River; about three-quarters of this eastern coal has greater than 1.5 lb S/MMBtu and some sort of treatment will be required to achieve environmental quality standards at coal-fired power plants.

The remaining third of total 1985 production is estimated to come from the West. About two-thirds of this western coal has less than 0.7 lbs S/MMBtu and could be used directly in power plants. Note, however, that the development outlook for this lowest sulfur coal is clouded by the litigation in the Sierra Club vs Morton (Kleppe) suit, and there is real doubt whether the amounts estimated here can be realized in the next decade. Should there be shortfalls in production from western deposits, then it seems likely that expanded eastern production (and treatment of the higher sulfur coal involved) would be necessary to meet electric utility industry requirements.

Coal Resources and Reserves: Utilization

Up to now, the discussion has been primarily on the quantity and quality of coal deposits in the ground. Certain bituminous coals have chemical and physical properties which make them valuable for special purposes, especially for manufacture of coke for reduction of iron ore to pig iron, chemical uses, and export. These coals are considered too valuable to be used for production of process steam or electric power generation.

Table 15 shows data on resources, reserve base, and recoverable reserves of bituminous coals in the United States with respect to coking properties. "Premium-and marginal-grade" coking coals[6] are distinguished from total bituminous coals in an attempt to arrive at an estimate of the magnitude of noncoking coals that would be of principal interest for steam-raising applications. The table shows that only about 11 billion tons of estimated recoverable reserves (16% of the total) are coking coals, and about 56 billion tons are estimated "steam" coals. Only the latter would appear to be available to meet electric utility industry requirements; because of the inability to substitute for coking coals in much of the steel production industry, it appears that most coking coals will have to be dedicated to metallurgical use. Because coking coals are low in sulfur content, the remaining coals for steam-raising must necessarily be higher in sulfur content. If these higher sulfur coals are to be used for power generation in compliance with air quality control standards, some sort of coal treatment, emission controls, or fuel/load switching (or combination of all approaches) would appear to be required. However, the fact that positive emission control technology remains to be proven reliable and economical in routine, continuous usage probably means that it will be essentially unavailable to meet power generation requirements much before the end of the century. This will place further emphasis on the search for coals with inherently low sulfur content or which can be simply processed or cleaned to satisfy emission control requirements.

Comparison of Projected Coal Supply and Demand for Electric Utility Use

Estimated future coal requirements for electric utilities in the contiguous United States were taken from data published by the National Electric Reliability Council (NERC).[7] These data give estimated total coal requirements by year to 1984; eastern coal requirements were calculated by the writers (Table 16).

Estimated future coal production for electric utility use was derived from data published by the National Coal Association (NCA).[8] These data give estimated total coal production, in addition to current production: It is anticipated that the end uses of this new production is expected to be similar to current production; about two-thirds of eastern coal will be dedicated to steam uses and about one-third to metallurgical uses, while essentially all western will be for steam. To derive estimated coal production for electric utility requirements, these factors were employed in re-calculation of the totals published by NCA.

The following assumptions were used in estimating future coal supply:

(a) The supply base of eastern steam coal was assumed to be 340 million tons. To allow for depletion of established production operations, this amount was reduced by 3% per year (the average historical depletion rate).

(b) The supply base of western steam coal was assumed to be 60 million tons. This amount was not reduced for depeletion, because it is "new" production and depletion is not anticipated in the next ten years.

(c) Planned or additional tonnages compiled by the NCA were reduced by 10 percent to allow for actual capacity utilization. Because of labor, maintenance, transportation, or other problems, a 90% capacity utilization factor is experienced.

The resulting estimates of new coal production were added to the supply base to obtain total estimated coal supply for electric utilities in the next decade (Table 16). The data are illustrated in Figure 2.

The table shows that deficits in coal supply are projected for eastern electric utilities beginning in 1979. These deficits increase to more than 130 million tons by 1984.

Although a surplus of supply over demand is shown for western utilities, the magnitude of the eastern deficit is such that an overall supply deficit results beginning in 1983.

To be sure, it is important to note NCA's caution that estimated coal supply additions are rather uncertain past 1981. This could reduce the size of estimated supply deficits, but it is not at all sure that the deficits would be easily eliminated. Note further that the situation indicated by the data in Table 1 is optimistic: projected coal supply from the East is not reduced because of sulfur content restrictions and western supplies are not curtailed by environmental litigation. In short, Table 1 presents minimum coal supply deficits. Actions that adversely impact supply conditions would both increase to size of the supply deficit and cause the deficit to occur earlier.

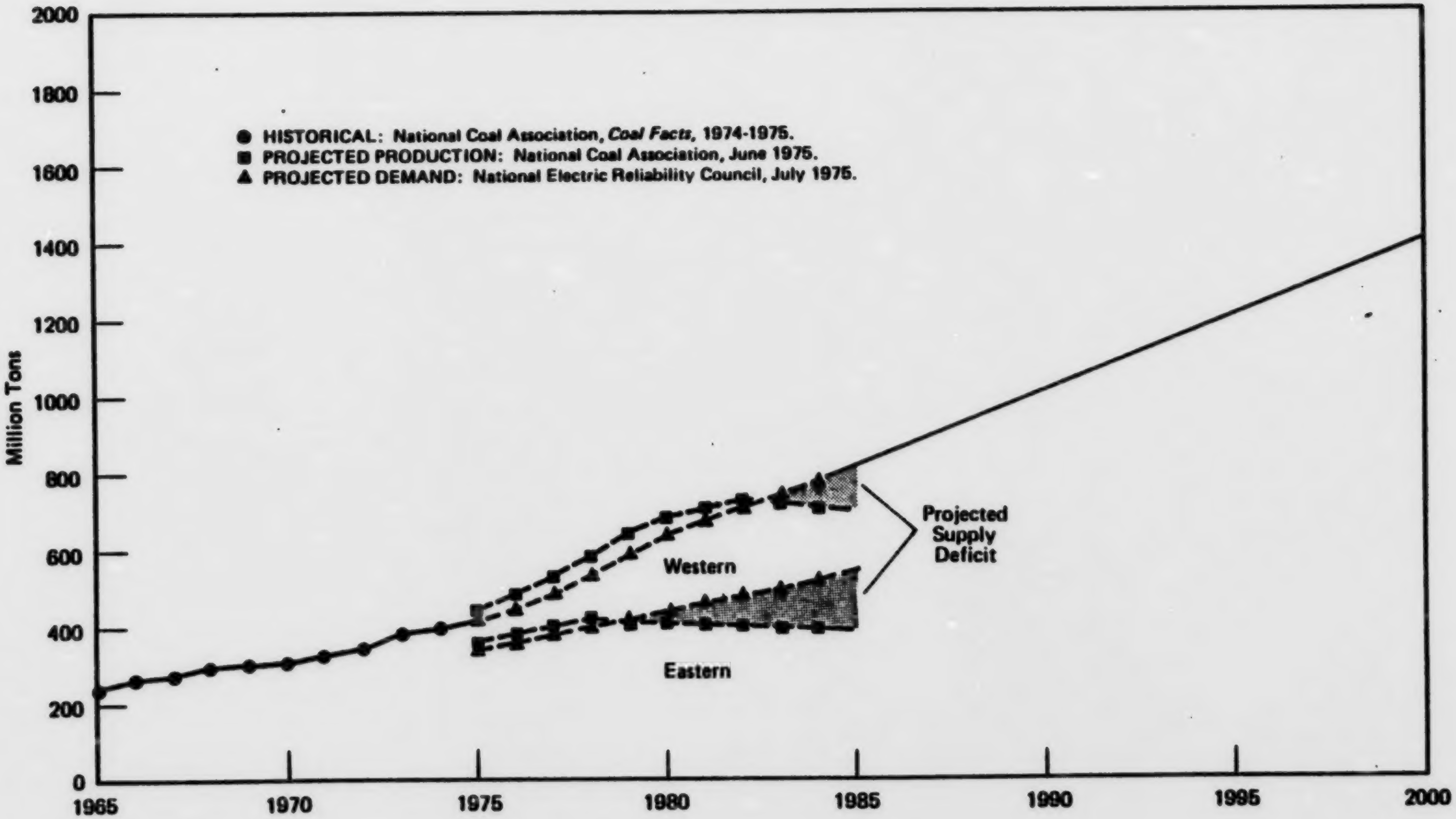

Figure 1 Comparison of Projected Coal Supply and Demand for Electric Utility Use

Also of concern is the apparent surplus of western coal supply over demand, which amounts to 60-70 million tons by the early 1980's. Western coal is of lower heating value and markedly different combustion conditions in contrast to eastern coal, and cannot be simply used as a replacement for shortfalls in eastern production. Therefore, while Table 1 suggests that the quantity of projected western coal supplies may help to meet indicated coal demand, the lower quality will lessen its contribution to reduction of the indicated supply deficit in the East. In this situation, the electric utilities would have the worst of both possible worlds -- a supply deficit in the East and overcapacity in coal production in the West that cannot really be employed effectively to eliminate the deficit.

Clearly, great care is required in matching coal production operations with electric utility demand so as to avoid the potential supply paradox indicated above.

SUMMARY

Recoverable reserves of coal are of the order of 8 percent of the total remaining resources. Natural geological conditions, as well as disturbances to strata from previous mining operations, may combine to make unminable coal seams which otherwise might be considered as potential reserves.

The quality of coal exerts an important influence on its ultimate use. The reserves of coking-type coals must be subtracted from the overall coal reserve figures, because of their dedication to the needs of the steel industry. Electric utility coal consumers are concerned about reliable supplies of coals having (at a minimum) satisfactory heating value and sulfur content. Uncertainties about factors critical to the development of coal production operations could have a serious adverse impact upon the availability of future coal supplies to meet projected demands.

REFERENCES

1. This estimate includes coals to a depth of about 3,000 feet. Additional coals are known to occur down to about 6,000 feet, but these are not considered further in this treatment because their extraction prospects are regarded as speculative.

2. U.S. Bureau of Mines, "Strippable Reserves of Bituminous Coal and Lignite in the United States," Information Circular 8531, 1971.

3. See, for example, Phelps, E.R. "Modern Mining Methods-Surface" in Elements of Practical Coal Mining," Cassidy, Editor, Am. Institute Mining, Metallurgical, and Petroleum Engineers, New York, 1973, page 377.

4. It is recognized that the reserve base estimates and the recoverable strippable reserves data were prepared at different times and using similar but somewhat different criteria. Nevertheless, the data are not significantly different, lending a measure of confidence that they could be employed together. This approach is intended to provide a rough approximation only.

5. Reiber, M., "Low Sulfur Coal: A Revision of Reserve and Supply Estimates," University of Illinois, Center for Advanced Computation, CAC Document No. 88, November 30, 1973.

6. "Premium-grade" coking coals have sulfur content of less than 1.3% and ash content of less than 8.1%. "Marginal-grade" coking coals have from 1.3 to 1.8% sulfur and ash from 8.1 to 12.0%.

7. National Electric Reliability Council, "Estimated Fossil Fuel Requirements, Projected Generating Capacity, and Electric Energy Production for the Electric Utility Industry (Contiguous U.S.) 1975-1984," Manuscript, July 1975.

National Coal Association, "New Coal Mines and Major Expansions of Existing Mines Planned, Announced, or Under Construction in the United States; 1975-1985," Manuscript, June 1975.

TABLE 1

SUMMARY OF RECOVERABLE COAL RESERVES BY RANK (MILLION TONS)

Rank	Strippable	Deep	Total	Percent of Total Resources
Anthracite	45	2,186	2,231	10
Bituminous	13,597	56,616	70,213	10
Subbituminous	24,318	29,282	53,600	13
Lignite	7,071	--	7,071	2
TOTAL	45,031	88,084	133,115	8

TABLE [illegible]

ANTHRACITE (million tons)	(1) Total Estimated Identified Resources Jan.1,1972 0-3000 ft.	(2) Total Measured and Indicated "Reserve Base" Jan.1,1974 0-1000 ft.	(3) Total Strippable Reserve Base 0-150 ft.	(4) Total Underground Reserve Base (2) - (3)	(5) Projected Recoverable Strippable Reserves (4) x 50%	(6) Projected Recoverable Underground Reserves (4) x 30%	(7) Total Projected Recoverable Reserves (5) + (6)	(8) Recoverable Resources as Percent of Identified Resources (7) ÷ (1)
ARKANSAS	430	96	--	96	--	28	28	7
COLORADO	78	28	--	28	--	8	8	10
NEW MEXICO	4	2	--	2	--	neg	neg	--
PENNSYLVANIA	20,510	7,120	90	7,030	45	2,109	2,154	11
VIRGINIA	335	138	--	138	--	41	41	12
WASHINGTON	5	--	--	--	--	--	--	--
TOTAL	21,362	7,384	90	7,294	45	2,186	2,231	10

Sources: Column (1): U.S. Geological Survey Professional Paper 820, p. 137

Columns (2) and (3): Project Independence Report, Coal Task Force, Appendix A

28 inch thickness = Anthracite & Bituminous

60 inch thickness = Subbituminous & Lignite

<1000 ft. depth

TABLE 3

BITUMINOUS (million tons)	(1) Total Estimated Identified Resources Jan.1,1972 0-3000 ft.	(2) Total Measured and Indicated "Reserve Jan.1,1974 0-1000 ft.	(3) Total Strippable Reserve Base 0-150 ft.	(4) Total Underground Reserve Base (2) - (3)	(5) Estimated strippable Reserves (USBM IC 8531)	(6) Projected Recoverable Reserves (4) x 30%	(7) Total Projected Recoverable (5) + (6)	(8) Recoverable Resources as Percent of Identified Resources (7) ÷ (1)
ALABAMA	13,342	1,955	157	1,798	134	539	673	5
ALASKA	19,413	1,201	1,201	--	480	--	480	2
ARIZONA	21,246	--	--	--	--	--	--	--
ARKANSAS	1,638	537	231	306	149	92	241	15
COLORADO	62,339	10,097	870	9,227	500	2,768	3,260	5
GEORGIA	24	1	--	--	--	--	--	--
ILLINOIS	139,124	65,665	12,223	53,442	3,247	16,032	19,279	14
INDIANA	34,573	10,623	1,674	8,949	1,096	2,685	3,781	11
IOWA	6,509	2,885	1,000	2,885	180	865	1,045	16
KANSAS	18,674	1,388	1,388	--	375	--	375	2
KENTUCKY	64,842	25,541	7,354	18,187	1,758	5,456	7,214	11
MARYLAND	1,158	1,048	146	902	21	271	292	25
MICHIGAN	205	119	1	118	1	35	36	18
MISSOURI	31,014	9,487	3,414	6,073	1,160	1,822	2,982	10
MONTANA	2,299	1,384	--	1,384	--	415	415	18
NEW MEXICO	10,752	1,777	250	1,527	--	458	458	4
N. CAROLINA	110	31	--	31	--	9	9	8
OHIO	41,358	21,077	3,654	17,423	1,033	5,277	6,260	15
OKLAHOMA	3,281	1,294	434	860	111	258	369	11
OREGON	50	--	--	--	--	--	--	--
PENNSYLVANIA	56,759	23,880	1,091	22,789	752	6,837	7,589	13

TABLE 3
(Concluded)

BITUMINOUS (million tons)	(1) Total Estimated Identified Resources Jan.1,1972 0-3000 ft.	(2) Total Measured and Indicated "Reserve Jan.1,1974 0-1000 ft.	(3) Total Strippable Reserve Base 0-150 ft.	(4) Total Underground Reserve Base (2) - (3)	(5) Estimated strippable Reserves (USBM IC 8531)	(6) Projected Recoverable Reserves (4) x 30%	(7) Total Projected Recoverable (5) + (6)	(8) Recoverable Resources as Percent of Identified Resources (7) ÷ (1)
TENNESSEE	2,572	987	320	667	74	200	274	11
TEXAS	6,048	--	--	--	--	--	--	--
UTAH	23,541	4,043	262	3,781	150	1,134	1,284	5
VIRGINIA	9,352	3,513	679	2,834	258	850	1,108	12
WASHINGTON	1,867	251	--	251	--	75	75	4
WEST VIRGINIA	100,628	39,590	8,984	30,606	2,118	9,181	11,299	11
WYOMING	12,705	4,524	--	4,524	--	1,357	1,357	11
OTHER	610	--	--	--	--	--	--	--
TOTAL	686,033	232,858	45,333	188,564	13,597	56,616	70,213	10

OTHERS = California, Idaho, Nebraska, Nevada

* No coal shown in strippable reserve base for Iowa in Project Independence. Value from IC 8531 used here.

+ Value shown in strippable reserve base for West Virginia in Project Independence is less than IC 8531 recoverable reserves. IC 8531 value of "recoverable strippable resource" (roughly equivalent to reserve base) used here instead.

TABLE 4

SUBBITUMINOUS	(2) Total Estimated Identified Resources Jan.1,1972 0-3000 ft	(3) Total Measured and Indicated "Reserve Base" Jan.1,1974 0-1000 ft	(4) Total Strippable Reserve Base 0-150 ft	(5) Total Underground Reserve Base (3) - (4)	(6) Estimated Recoverable Strippable Reserves (USBM IC 8531)	(7) Projected Recoverable Underground Reserves (5) x 30%	(8) Total Projected Recoverable Reserves (6) + (7)	(9) Recoverable Resources as Percent of Identified (8) ÷ (2)
ALASKA	110,668	10,148	5,902	4,246	3,926	1,274	5,200	5
ARIZONA	--	350	400*	--	387	--	387	
COLORADO	18,242	4,745	--	4,745	--	1,424	1,424	8
MONTANA	131,855	99,212	35,431	63,781	3,400	19,134	22,534	17
NEW MEXICO	50,671	2,615	2,646+	607	2,474	182	2,656	5
OREGON	284	1	neg	--	--	--	--	--
UTAH	180	--	--	--	--	--	--	--
WASHINGTON	4,190	1,695	500	1,195	135	359	494	12
WYOMING	107,951	46,704	23,674	23,030	13,971	6,909	20,880	19
OTHERS	32	--	--	--	--	--	--	--
TOTAL	424,073	165,470	67,865	97,604	24,318	29,282	53,575	13

OTHERS= California, Idaho

*Project Independence reserve base for Arizona less than IC 8531 recoverable reserves; IC 8531 data used here.

+IC 8531 data used here.

TABLE 5

LIGNITE	(2) Total Estimated Identified Resources Jan.1,1972 0-3000 ft.	(3) Total Measured and Indicated "Reserve Base" Jan.1,1974 0-1000 ft.	(4) Total Strippable Reserve Base 0-250 ft.	(5) Total Underground Reserve Base (3) - (4)	(6) Estimated Recoverable Strippable Reserves (USBM IC 8531)	(7) Projected Recoverable Underground Reserves (5) x 30%	(8) Total Projected Recoverable Reserves (6) + (7)	(9) Recoverable Reserves as Percent of Identified Resources (8) ÷ (2)
ALABAMA	2,000	1,027	1,027	--	--	--	--	--
ALASKA	(?)	296	--	--	--	--	--	--
ARKANSAS	350	32	--	--	25	--	25	7
MONTANA	87,521	7,131	7,131	--	3,497	--	3,497	4
N. DAKOTA	350,630	16,003	16,003	--	2,075	--	2,075	1
S. DAKOTA	2,031	428	428	--	160	--	160	8
TEXAS	6,824	3,272	3,272	--	1,309	--	1,309	19
WASHINGTON	117	8	8	--	--	--	--	--
OTHERS	46	--	--	--	--	--	--	--
TOTAL	449,519	28,197	27,901	0	7,071	0	7,071	2

OTHERS = California, Idaho, Louisiana, Mississippi

Notes to Tables 2-5

(1) Source is U.S. Bureau of Mines, "Demonstrated Coal Reserve Base of the United States on January 1, 1974," Mineral Industry Survey, June 1974.

Note that the maximum depth for all ranks of coal except lignite, is 1000 feet. Only the lignite beds that can be mined by surface methods are included; depths are less than about 120 feet. Seams greater than the following thicknesses are included: bituminous coal and anthracite, 28 inches or more; subbituminous coal and lignite, 60 inches or more.

Note that these data include 100% of the coal in-place, and no recoverability factor has been included.

(2) Source is EPRI, using data from the U.S. Bureau of Mines and U.S. Geological Survey.

(3) Source is U.S. Bureau of Mines, "Strippable Reserves of Coal and Lignite in the United States," Information Circular 8531, 1971.

TABLE 6

INDICATED RECOVERY PERCENTAGE FOR STRIPPABLE DEPOSITS (RESERVE BASE + RECOVERABLE RESERVES)

STATE	ANTHRACITE	BITUMINOUS	SUBBITUMINOUS	LIGNITE
ALABAMA	--	85	--	--
ALASKA	--	40	67	--
ARIZONA	--	--	97	--
ARKANSAS	--	65	--	78
COLORADO	--	57	--	--
GEORGIA	--	--	--	--
ILLINOIS	--	27	--	--
INDIANA	--	65	--	--
IOWA	--	18	--	--
KANSAS	--	27	--	--
KENTUCKY	--	24	--	--
MARYLAND	--	14	--	--
MICHIGAN	--	100	--	--
MISSOURI	--	34	--	--
MONTANA	--	--	10	49
NEW MEXICO	--	--	93	--
NO. CAROLINA	--	--	--	--
NO. DAKOTA	--	--	--	13
OHIO	--	28	--	--
OKLAHOMA	--	26	--	--
OREGON	--	--	--	--
PENNSYLVANIA	76	69	--	--
SO. DAKOTA	--	--	--	37
TENNESSEE	--	23	--	--
TEXAS	--	--	--	40
UTAH	--	57	--	--
VIRGINIA	--	38	--	--
WASHINGTON	--	--	27	--
WEST VIRGINIA	--	24	--	--
WYOMING	--	--	59	--
OTHERS	--	--	--	--
MEAN		41	49	43

TABLE 7

SUMMARY COMPARISON OF ESTIMATES OF COAL RESOURCES AND RESERVES ($\leq$0.7% SULFUR) JANUARY 1, 1965

(Million Short Tons)

	Conventional Estimates (Tonnage Only)		Standardized Estimates (Normalized for Btu/Sulfur)	
	Resources	Reserves	Resources	Reserves
Appalachian				
Bituminous	37,320	4,105	44,784	4,926
Anthracite	12,550	1,630	14,056	1,826
Interior				
Bituminous	445	70	472	74
Rockies				
Bituminous	45,215	4,585	48,581	4,925
Subbituminous	181,670	19,470	46,329	4,632
Lignite	344,620	37,905	0	0
West Coast				
Bituminous	900	80	855	76
Subbituminous	3,780	340	0	0
Total	626,500	68,185	155,077	16,459

TABLE 8

POTENTIAL RECOVERABLE RESERVES OF ANTHRACITE (MILLION TONS)

	(1) RECOVERABLE STRIPPABLE RESERVES	(2) RECOVERABLE UNDERGROUND RESERVES	(3) TOTAL RECOVERABLE RESERVES	(4) HEATING VALUE MM Btu/ton	(5) Lbs S MM btu	(6) REMARKS	(7) ASSESSMENT OF POTENTIAL RECOVERABLE RESERVES
ARKANSAS	--	28	28	28.0	1.64	S too high; reserves low; development unlikely	0
COLORADO	--	8	8	27.9	0.43	S satisfactory; reserves low; development unlikely	0
NEW MEXICO	--	neg	neg	27.2	0.66	S marginal; reserves low; development unlikely	0
PENNSYLVANIA	45	2,109	2,154	26.2	0.51	S satisfactory; reserves OK development likely	2,154
VIRGINIA	--	41	41	23.8	0.59	S marginal; reserves low; development unlikely	0
WASHINGTON	--	--	--	--	--		--
TOTAL	45	2,186	2,231				2,154

ASSESSED RESERVES < 0.7 lb S/MM Btu = 2,154 million tons

> 0.7 lb S/MM Btu = 0

TOTAL 2,154

TABLE 9

POTENTIAL RECOVERABLE RESERVES OF BITUMINOUS COAL (MILLION TONS)

	(1) RECOVERABLE STRIPPABLE RESERVES	(2) RECOVERABLE UNDERGROUND RESERVES	(3) TOTAL RECOVERABLE RESERVES	(4) HEATING VALUE MM Btu/ton	(5) Lbs S MM Btu	(6) REMARKS	(7) ASSESSMENT OF POTENTIAL RECOVERABLE RESERVES
ALABAMA	134	539	673	27.6	0.80	S medium; reserves adequate; development likely	673
ALASKA	480	--	480	24.6	0.16	S low; reserves adequate; development likely	480
ARIZONA	--	--	--	23.2	0.78	S medium; reserves low; development unlikely	0
ARKANSAS	149	92	241	28.2	1.60	S high; reserves marginal; development unlikely	0
COLORADO	500	3,768	3,268	25.6	0.55	S satisfactory; reserves high; development likely	3,268
GEORGIA	--	--	--	--	--	--	0
ILLINOIS	3,247	16,032	19,279	25.2	2.54	S high; reserves large; development possible	19,279
INDIANA	1,096	2,685	3,781	26.0	2.31	S high; reserves large; development possible	3,781
IOWA	180	865	1,045	23.4	4.70	S high; reserves adequate; development unlikely	0
KANSAS	375	--	375	25.4	2.38	S high; reserves marginal; development possible	375
KENTUCKY	1,758	5,456	7,214	26.8	1.87	S high; reserves large; development possible	7,214
MARYLAND	21	271	292	28.2	1.00	S medium; reserves marginal; development likely	292
MICHIGAN	1	35	36	26.2	1.54	S high; reserves small; development unlikely	0
MISSOURI	1,160	1,822	2,982	24.6	3.58	S high; reserves large; development unlikely	0
MONTANA	--	415	415	24.4	1.07	S medium; reserves adequate; development likely	415
NEW MEXICO	--	458	458	25.8	0.54	S low; reserve adequate; development likely	458

TABLE 9
(Concluded)

	(1) RECOVERABLE STRIPPABLE RESERVES	(2) RECOVERABLE UNDERGROUND RESERVES	(3) TOTAL RECOVERABLE RESERVES	(4) HEATING VALUE MM Btu/ton	(5) Lbs S MM Btu	(6) REMARKS	(7) ASSESSMENT OF POTENTIAL RECOVERABLE RESERVES
NO. CAROLINA	--	--	--	--	--	--	0
OHIO	1,033	5,227	6,260	25.8	2.25	S high; reserves large; development possible	6,260
OKLAHOMA	111	258	369	27.6	1.52	S high; reserves marginal; development possible	369
OREGON	--	--	--	22.4	0.63	S low; reserves negligible; development unlikely	0
PENNSYLVANIA	752	6,837	7,589	27.6	1.31	S medium; reserves large; development likely	7,589
TENNESSEE	74	200	274	27.2	1.03	S medium; reserves marginal; development likely	274
TEXAS	--	--	--	25.8	1.78	S high; reserves negligible; development unlikely	0
UTAH	150	1,134	1,284	26.8	0.37	S low; reserves adequate; development likely	1,284
VIRGINIA	258	850	1,108	28.0	0.57	S low; reserves adequate; development likely	1,108
WASHINGTON	--	75	75	24.8	0.48	S low; reserves small; development possible	75
WEST VIRGINIA	2,118	9,181	11,299	27.4	1.82	S high; reserves large; development possible	11,299
WYOMING	--	1,357	1,357	25.8	0.54	S low; reserves adequate; development likely	1,357
OTHER	--	--	--	--	--	--	--
TOTAL	13,597	56,616	70,213				65,850

ASSESSED RESERVES
< 0.7 lb S/MM Btu = 8,030 million tons
> 0.7 < 1.5 = 9,238
> 1.6 < 2.1 =15,262
> 2.2 =29,695

Likely = 17,193
Possible = 45,032
Unlikely = 4,304

TABLE 10

POTENTIAL RECOVERABLE RESERVES OF SUBBITUMINOUS COAL (MILLION TONS)

	(1) RECOVERABLE STRIPPABLE RESERVES	(2) RECOVERABLE UNDERGROUND RESERVES	(3) TOTAL RECOVERABLE RESERVES	(4) HEATING VALUE MM Btu/ton	(5) Lbs S MM BTU	(6) REMARKS	(7) ASSESSMENT OF POTENTIAL RECOVERABLE RESERVES
ALASKA	3,926	1,274	5,200	22.4	0.18	S low; reserves large; development possible	5,200
ARIZONA	387	--	387	23.2	0.78	S marginal; reserves adequate; development possible	387
COLORADO	--	1.424	1,424	24.6	0.33	S low; reserves adequate; development likely	1,424
MONTANA	3,400	19,134	22,534	23.4	0.95	S marginal; reserves large: development possible	22,534
NEW MEXICO	2,474	182	2,656	23.0	1.05	S medium; reserves adequate; development possible	2,656
OREGON	--	--	--	22.4	0.63	S low; reserves negligible; development unlikely	0
UTAH	--	--	--	--	--	--	0
WASHINGTON	135	359	494	22.2	0.72	S medium; reserves adequate; development possible	494
WYOMING	13,971	6,909	20,880	24.6	0.49	S low; reserves large; development likely	20,880
OTHERS	--	--	--			--	--
TOTAL	24,318	29,282	53,575			Likely = 22,304 Possible = 31,271	53,575

ASSESSED RESERVES < 0.7 lb S/MM Btu = 27,504 million tons

> 0.7 < 1.5 = 26,071

TABLE 11

POTENTIAL RECOVERABLE RESERVES OF LIGNITE (MILLION TONS)

	(1) RECOVERABLE STRIPPABLE RESERVES	(2) RECOVERABLE UNDERGROUND RESERVES	(3) TOTAL RECOVERABLE RESERVES	(4) HEATING VALUE MM Btu/ton	(5) Lbs S MM Btu	(6) REMARKS	(7) ASSESSMENT OF POTENTIAL RECOVERABLE RESERVES
ALABAMA	--	--	--	--	--	--	0
ARKANSAS	25	--	25	--	--	--	0
MONTANA	3,497	--	3,497	22.4	0.54	S low; reserves large; development likely	3,497
N. DAKOTA	2,075	--	2,075	21.8	0.64	S low; reserves large; development likely	2,075
S. DAKOTA	160	--	160	21.8	0.92	S marginal; reserves small; development unlikely	0
TEXAS	1,309	--	1,309	22.0	1.27	S medium; reserves adequate; development possible	1,309
WASHINGTON	--	--	--	--	--	--	0
OTHERS	--	--	--	--	--	--	0
TOTAL	7,071	0	7,071			Likely = 5,572 Possible= 1,309	6,881

ASSESSED RESERVES < 0.7 lb S/MM Btu = 5,572 million tons
> 0.7 < 1.5 = 1,309

TABLE 12

SUMMARY OF POTENTIAL RECOVERABLE RESERVES (MILLION TONS)

RANK	<0.7 lbs/MMBtu	>0.7 <1.5 lbs/MMBtu	>1.5 <2.1 lbs/MMBtu	>2.1 lbs/MMBtu	Total
ANTHRACITE	2,154	—	—	—	2,154
BITUMINOUS	8,030	9,238	15,262	33,320	65,850
SUBBITUMINOUS	27,504	26,071	—	—	53,440
LIGNITE	5,572	1,309	—	—	6,881
TOTAL	43,260	36,618	15,262	33,320	128,325

TABLE 13

COMPARISON OF ASSESSED RECOVERABLE RESERVES OF BITUMINOUS AND SUBBITUMINOUS COAL
AGAINST PROJECTED PRODUCTION (MILLION TONS)

	(1)	(2)	(3)	(4)	(5)	(6)	(7)			
	ASSESSED RECOVERABLE RESERVES			LBS S MM BTU	PRESENT PRODUCTION (1973)	FORECAST NEW CAPACITY (by 1985)	OPTIMISTIC PRODUCTION IN 1985 (5) + (6)	OPTIMISTIC PRODUCTION BY SULFUR CONTENT (Lbs/MM Btu)		
	LIKELY	POSSIBLE	UNLIKELY					<0.7	>0.7<1.5	>1.5
ALABAMA	673			0.80	19.2	19.6	34.0		34.0	
ALASKA	480	5,200(s)		0.16	0.7		1.0	1.0	1.0	
ARIZONA		387	neg	0.78	3.3	27.0	29.5		29.5	
ARKANSAS			neg	1.60	0.4		0.5			0.5
COLORADO	1,424(s) 3,268			0.55	6.2	2.9	7.6	7.6		
GEORGIA			neg							
ILLINOIS		19,279		2.54	61.6	40.6	86.8			86.8
INDIANA		3,781		2.31	25.3	11.0	30.0			30.0
IOWA			1,045	4.70	0.6		1.0			1.0
KANSAS		375		2.38	1.1		1.0			1.0
KENTUCKY		7,214		1.87	127.7	69.0	164.8			164.8
MARYLAND	292			1.00	1.8		2.0		2.0	
MICHIGAN			36	1.54						
MISSOURI			2,982	3.58	4.7		5.0			5.0
MONTANA	415	22,534(s)		1.07	10.4	57.3	65.1		65.1	
NEW MEXICO	458	2,656(s)		0.54	9.1	13.6	20.4	20.4		
N. CAROLINA			neg							
OHIO		6,260		2.25	45.8	14.2	48.6			48.6
OKLAHOMA		369		1.52	2.2		2.0			2.0
OREGON			neg	0.63						

TABLE 13
(Concluded)

	(1)	(2)	(3)	(4)	(5)	(6)	(7)			
	ASSESSED RECOVERABLE RESERVES							OPTIMISTIC PRODUCTION BY SULFUR CONTENT (Lbs/MM Btu)		
	LIKELY	POSSIBLE	UNLIKELY	LBS S MM BTU	PRESENT PRODUCTION (1973)	FORECAST NEW CAPACITY (by 1985)	OPTIMISTIC PRODUCTION IN 1985 (5) + (6)	<0.7	>0.7<1.5	>1.5
PENNSYLVANIA	7,589			1.31	76.4	31.5	88.8		88.8	
TENNESSEE	274			1.03	8.2	4.9	11.1		11.1	
TEXAS			neg	1.78						
UTAH	1,284		neg(s)	0.37	5.5	36.6	40.7	40.7		
VIRGINIA	1,108			0.57	34.0	13.7	39.2	39.2		
WASHINGTON		494(s) 75		0.48	3.3	3.0	5.5	5.5		
WEST VIRGINIA		11,299		1.82	115.5	71.0	157.6			157.6
WYOMING	20,880(s) 1,357			0.54	14.9	123.4	134.6	134.6		
OTHER			neg							
TOTAL	17,193 22,034(s)	48,657 31,271(s)	4,304 neg(s)		577.9	539.3	976.8	249.0	230.5	497.3

TABLE 14

COMPARISON OF ASSESSED RECOVERABLE RESERVES OF LIGNITE AGAINST PROJECTED PRODUCTION (MILLION TONS)

	ASSESSED RECOVERABLE RESERVES			LBS S MM BTU	PRESENT PRODUCTION (1973)	FORECAST NEW CAPACITY (to 1985)	OPTIMISTIC PRODUCTION IN 1985*	OPTIMISTIC PRODUCTION BY SULFUR CONTENT (Lbs/MM Btu)		
	LIKELY	POSSIBLE	UNLIKELY					<0.7	>0.7 <1.5	>1.5
ALABAMA			neg	--	--					
ARKANSAS			neg	--	--					
MONTANA	3,497			0.54	0.3		0.3	0.3		
N. DAKOTA	2,075			0.64	6.9	21.9	27.1	27.1		
S. DAKOTA			160	0.92	--	--	--	--		
TEXAS		1,309		1.27	6.9	16.7	21.9		21.9	
WASHINGTON				--	--					
OTHERS				--	--					
TOTAL	5,572	1,309	160		14.1	38.6	49.3	27.4	21.9	0

* See note on Table 13.

TABLE 15

Resources, Reserve Base, And Recoverable Reserves of Bituminous Coals in the United States with Respect to Coking Properties
(billion tons)

Item	(1) Total Bituminous Coal	(2) Coking Coals Premium-Grade	(3) Coking Coals Marginal Grade	(4) Coking Coals Total	(5) Estimated "Steam" Coal (1)-(4)
1. Remaining Identified Resources	747(a)	101(b)	82(b)	183(b)	564
2. Demonstrated Reserve Base	219(c)	20(d)	16(e)	36	183
3. Estimated Recoverable Reserves	67(f)	6(g)	5(g)	11	56

Notes

(a) Averitt, P. "Coal Resources of the United States, January 1, 1974," U.S. Geological Survey, Bulletin 1412, 1975.

(b) Sheridan, E. T. "United States Coals for Coke Production," Presented at Symposium on Developments in European and World Markets for Coking Coal and Coke," Rome, March 20-30, 1973 (Manuscript).

(c) U.S. Bureau of Mines, "Demonstrated Coal Reserve Base of the United States on January 1, 1974," Mineral Industry Surveys, June 1974.

(d) Mutschler, P. H., "Impact of Changing Technology on the Demand for Metallurgical Coal and Coke Produced in the United States to 1985," U.S. Bureau of Mines Information Circular No. 8677, 1975.

(e) Estimated at same proportion of resources as premium-grade coking coals.

(f) This report, Table _______

(g) Calculated assuming 30 percent recovery from reserve base (see preceding section).

TABLE 16

COMPARISON OF PROJECTED COAL SUPPLY AND DEMAND FOR ELECTRIC UTILITY USE

	1975	1976	1977	1978	1979	1980	1981	1982	1983	1984
EAST										
Supply: (a)	340.0	329.8	319.6	309.4	299.2	289.0	278.8	268.6	258.4	248.2
(b)	17.9	45.7	71.0	96.7	116.5	129.3	136.6	141.6	145.2	148.6
Total Supply	357.9	375.5	390.6	406.1	415.7	418.3	415.4	410.2	403.6	396.8
Demand: (NERC)	345.2	365.7	381.0	405.2	419.2	444.7	461.2	479.1	499.8	528.1
Surplus of Supply	12.7	9.8	9.6	0.9	(3.5)	(26.4)	(45.8)	(68.9)	(96.2)	(131.3)
WEST										
Supply: (c)	60.0	60.0	60.0	60.0	60.0	60.0	60.0	60.0	60.0	60.0
(b)	19.5	49.5	80.2	119.4	171.5	208.5	226.7	242.8	250.0	253.6
Total Supply	79.5	104.5	140.2	179.4	231.5	268.5	286.7	302.8	310.0	313.6
Demand: (NERC)	74.4	88.0	106.8	135.5	171.6	196.3	213.2	231.8	240.8	252.7
Surplus of Supply	5.1	21.5	33.4	43.9	59.9	72.2	73.5	71.0	69.2	60.9
TOTAL SUPPLY: (d)	400.0	389.8	379.6	369.4	359.2	349.0	338.8	328.6	318.4	308.2
(b)	37.4	95.2	151.2	216.1	288.0	337.8	363.3	384.4	395.2	402.2
	437.4	485.0	530.8	585.5	647.2	686.8	702.1	713.0	713.6	710.4
DEMAND: (NERC)	419.6	453.7	487.8	540.7	590.8	641.0	674.4	710.9	740.6	780.8
Surplus of Supply	17.8	31.3	43.0	44.8	56.4	45.8	27.7	2.1	(27.0)	(70.4)

(a) Supply Base - 340 million tons discounted 3%/year

(b) Source - NCA study on additional mine capacity, at 90% capacity factor

(c) Supply Base - 60 million tons

(d) Supply Base - addition of eastern and western

PROBLEMS AND CONTROL OPTIONS USING LOW SULFUR COAL IN UTILITY BOILERS

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PROBLEMS AND CONTROL OPTIONS
USING LOW SULFUR COAL IN UTILITY BOILERS

Presented by:

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Denver, Colorado

This paper describes briefly the problems associated with the handling and burning of low sulfur fuels and the control options available for flue gas emissions.

Aside from the question of emission control options, one must also consider the other problems associated with low sulfur, low rank western coals such as the handling of the coal, i.e., railroad cars and unloading facilities at the power plant. The paper will also point out the considerations to be given in design for pulverizers, increased primary air for pulverizer drying, and boiler thermal efficiencies. Additional considerations must be given to the boiler design to account for the particular flame characteristics of the low rank fuels along with the high fusion ash and the basic composition of the ash.

The paper will also present in chronological order the efforts of Public Service Company of Colorado (PSCo) in installing control equipment for the low sulfur western coals. A brief history of the PSCo experience with electrostatic precipitators to include the installation of cold side precipitators, the addition of gas conditioning, and, finally, the application of hot side electrostatic precipitators. It will also discuss the use of wet stack gas scrubbers for particulate collection and briefly review the use of bag filters.

Adequate particulate removal efficiencies can be obtained with all the control devices mentioned above, however, high operational and maintenance cost and, in some cases, a lower than satisfactory availability record for the devices must be considered in a case-by-case choice for each specific coal and location.

There is no one specific device that is recommended over and above the other control devices; it is a matter of judgement and good engineering practices on behalf of the owner and operator.

PROBLEMS AND CONTROL OPTIONS
USING LOW SULFUR COAL IN UTILITY BOILERS

INTRODUCTION

With the advent of the Clean Air Act and subsequent regulations promulgated by the Environmental Protection Agency, a great deal of emphasis has been placed on burning low sulfur western sub-bituminous coal. The object of this paper is to attempt to shed some light on the problems incurred with burning such coals. It will be of particular interest to note these problems when considering the conversion of present boilers now burning an eastern coal to a configuration to burn western low sulfur coals. Not only are there serious problems choosing the proper particulate control device, but there are problems to be considered in the handling of the western coals and boiler designs.

It is not my attempt to layout guidelines for the design and production of a new power generation facility but to dwell more on the problems that have occurred and that should be understood when considering a switch in fuel supply to the western low sulfur sub-bituminous coals. Since the promulgation of the Environmental Protection Agency regulations, several companies have attempted to convert present boilers, all of which have experienced some difficulty not only in the operation of the boiler but, probably more importantly, in the operation of the control devices associated with the boiler.

The Public Service Company of Colorado has been burning low sulfur coal since the 1920s and instituted the use of electrostatic precipitators in the early 1960s. The information contained herein is related to the experiences of the Public Service Company of Colorado and other companies which have installed systems using western coal as the main source of fuel.

WESTERN COAL

Western coal reserves, which are predominantly lower rank coals and include lignite, sub-bituminous and bituminous, have been measured by the United States Bureau of Mines to be 216 billion tons, as indicated in Table 1. The distribution of this coal is primarily in Montana, Wyoming, North Dakota and Colorado. During 1974 the estimated rate of production was greater than 2 million tons per year in each of the eleven western states, with Wyoming producing the greatest tonnage. It

TABLE 1 COAL IN THE WESTERN U. S., RESERVES AND PRODUCTION BY STATE

State	In-place coal reserve (In Million Tons)	Estimated 1974 production (In Million Tons)	New mine* capacity by 1983 (In Million Tons)
Arizona	350	3.2	8
Colorado	14,870	6.9	** NA
Kansas	1,388	.8	.5
Missouri	9,488	4.3	NA
Montana	107,727	13.6	21
New Mexico	4,394	9.5	5
North Dakota	16,003	7.2	NA
Oklahoma	1,294	2.4	NA
South Dakota	428	.0	NA
Texas	3,272	6.0	.1
Utah	4,042	6.5	1.3
Washington	1,954	3.9	1
Wyoming	51,228	20.5	82.5
	216,439	84.8	119.4

* New capacities reported by 1983 are representative of firm plans announced by major producing firm. The values given do not reflect plans for coal gasification.

** NA indicates that no value was given in the reference. Some of the states affected are known to be experiencing major expansions in coal production.

Sondreal and Tufte, 1975.

is anticipated that both Wyoming and Montana will experience the greatest increase in production of the western coals. As indicated in Table 1, plans for expansion of mines in the West will increase the production to over 200 million tons per year by 1983.

Important properties of western coals compared to eastern and midwestern coals include a lower sulfur content, generally averaging approximately 0.7 percent, a lower heating value, a higher moisture content, and higher alkali oxides such as Na_2O, MgO and CaO. "Alkali content tends to be the highest in the lowest rank coal, lignite, and progressively less prevalent in the sub-bituminous and bituminous coals. This trend is related to the greater ion exchange capacity of low-rank coals compared with the higher rank. Variations in alkali content are also influenced by the minerology of the overburden and the course of ground water movement. Alkali content in western coal ash varies from under 10 to over 50 pct, with important variations occurring within individual mines.

"A guideline for assessing the importance of the amount of alkali in western coal is the ratio of the alkali to coal sulfur. For a coal containing 7.5 pct ash and 20 pct alkali in the ash, the total alkali is chemically equivalent to slightly more than 120 pct of a 0.7 pct sulfur content. For some lignites, the alkali/sulfur ratio can be several hundred percent. Thus there is ample alkali to interact importantly with sulfur oxides in a wet scrubber in burning many western coals." (Sondreal and Tufte, 1975.)

For relative comparison purposes only, Table 2 shows typical analyses of western and midwestern coals. These do not represent any specific mines.

Transportation. At present, the majority of the western coals are strip mined and used at mine mouth-plant operations or transported to load centers by unit trains. In the location of the more recent large units there seems to be a preference for the use of the unit train concept.

In considering unit train operation, there are problems associated with the transportation of western coals which should be considered. For example, the lower rank coals are generally much higher in moisture content, so built into transportation cost is the added expense of hauling more moisture from the mine to the power plant. More tonnage is hauled because of the lower BTU, thus adding the expense of additional railroad cars.

In unit train operation, the cars travel approximately 100 thousand miles per year. In many cases, coal is loaded at the mine on the fly and unloaded at the power plant, also on the fly. Therefore, the cars virtually never come to a standstill except for momentary stoppage of the trains to transfer crews. This

TABLE 2 TYPICAL COAL AND ASH ANALYSIS

	Colorado Coal	Wyoming Coal	Midwestern Steam Coal[1]
Proximate Analysis as Received:			
Moisture	9.8%	29.0%	2.2%
Volatile Matter	35.3%	33.4%	35.6%
Fixed Carbon	45.5%	32.4%	53.3%
Ash	9.4%	5.2%	8.9%
Sulfur	0.7%	0.6%	2.2%
HHV, BTU/lb.	11,010	8,250	13,280
Hardgrove Grindability	44	55	-
Pounds of Sulfur/million BTU's	0.64	0.73	1.66
Ash Fusion Temperatures:			
Initial Deformation	2560°F.	2010°F.	2090°F.
Softening (H=W)	2665°F.	2150°F.	2210°F.
Fluid	2690°F.	2210°F.	2330°F.
Ash Composition:			
Silica (SiO_2)	54.1%	34.5%	47.2%
Alumina (AL_2O_3)	28.3%	20.6%	23.2%
Iron Oxide (Fe_2O_3)	3.0%	6.8%	21.9%
Phosphorus Pentoxide (P_2O_5)	1.1%	0.2%	.2%
Titanium Oxide (TiO_2)	0.7%	Trace	1.0%
Calcium Oxide (CaO)	5.0%	20.0%	2.4%
Magnesium Oxide (MgO)	1.5%	2.7%	.6%
Sodium Oxide (Na_2O)	0.6%	0.2%	.4%
Potassium Oxide (K_2O)	1.0%	4.3%	1.6%
Sulfur Trioxide (SO_3)	1.8%	10.7%	1.6%

1/ Pittsburgh No. 8 steam coal from Jefferson County, Ohio.

Green, G. P. and W. S. Landers, 1974.

factor becomes very important when considering that an average box car will travel approximately 17 thousand miles per year. Maintenance of rolling stock now takes on a new perspective; and when considering historical maintenance cost on an annual basis, it must now be increased several fold. In the past, a typical piece of rolling stock, traveling approximately 17 thousand miles per year, had an average life of twenty years, now it covers the equivalent mileage in three or four years.

Not only is there now greater annual wear in the moving parts of the car such as the wheels and truck assemblies, but the very nature of the western coal tends to aggrevate the problems of spontaneous combustion, thus necessitating additional care in providing tight railroad cars. Also, western coals are less dense than an eastern coal, so that when a car is fully loaded, the center of gravity will tend to shift upward and, in some cases, will induce a serious swaying in the train necessitating lower tonnage hauled per car than originally contemplated.

Since most unit trains are designed with special features such as swivel couplers or automatic dump hoppers, it is imperative that when entering into the unit train operation the utility company give serious consideration to how many spare railroad cars should be purchased to complement the original unit train. Actual practice has shown that a greater number of spare cars are necessary than were initially projected during the design of the unit train operation.

Handling. There appears to be a greater number of fines in the western coal. While in transit the fines tend to dry out rapidly causing a dust problem at unloading facilities. The dust problem can be minimized with the proper design and utilization of dust supression systems and wetting agents.

A greater percentage of fines also increases dust problems associated with long-term storage. In spite of efforts to compact coal in the storage piles, the inherent strong winds in the western portion of the United States tend to cause fugitive dust problems in the coal storage area. When designing long-term coal storage areas, special consideration has to be given to such items as lowering wells and bulk handling equipment. When handling the coal in storage, coal should not be allowed to fall any appreciable distance in the open air as it aggrevates the fugitive dust problem. It is highly recommended that such items as telescoping spouts be used to minimize dust associated with the fines.

Spontaneous combustion in coal piles tends to be greater with the western coal. Extensive surveillance by plant operation personnel as well as good practices in compacting stored coal are necessary. There is an economic consideration to the amount of compaction that can be obtained in the coal storage pile. This cost of compaction must offset against the increased surveillance for coal pile fires, and the individual utility company must determine which course of action is to its particular economic advantage.

BOILER OPERATING CONSIDERATION

Because of the higher inherent moisture and lower BTU of western coal, special consideration must be given to the design of the boiler and associated pulverizers and air preheaters. Larger air heaters are required, particularly for primary air heating, to vaporize the inherent moisture in the coal. By the same token, pre-heated air to the coal pulverizer cannot be of such a temperature to cause ignition of the coal in the pulverizer. In many cases, there is a fine point in determining the optimum mill temperature in relation to fires and explosions within the mills.

Since western coals have a lower heating value, greater tonnages are required, thereby increasing the size of the mills and increasing the capacity of the primary air fans. Lower boiler thermal efficiencies must be expected due to the loss associated with water vaporization, and this has a tendency to lower the fire box temperatures and requires greater heat absorption surfaces. In new designs, these factors are taken into consideration and can be compensated for. Greater problems exist in attempting to substitute the western coal in an existing installation which was designed for use with higher ranking coals.

Burning characteristics are quite different between the higher and lower rank coals. The lower rank coal (after mill drying) will ignite closer to the burner and at a faster rate, thus effecting the pattern of heat release and the location of additional heat absorption surface within the radiant section of the boiler.

Perhaps the most serious consideration to be given to the use of the western coals is their characteristic fusion temperatures. The high fusion temperature will generally not permit their use in wet bottom or slag tap boilers. As mentioned previously, the higher alkalinity in the ash content of the coal will also have an effect on the ash collection systems. Many boilers designed for the higher rank coals will encounter plugging in reheat and super-heat sections of the boiler when burning western coals.

CONTROL OPTIONS

The most difficult problem to overcome in the burning of low sulfur western fuels is the collection of particulate matter in the flue gas.

Mechanical Collectors. Initially, only mechanical collectors were utilized for the collection of particulate matter. Mechanical collectors which were properly designed and maintained exhibited an efficiency of approximately 80-85 percent removal, which by today's standards would be totally unacceptable.

Electrostatic Precipitators. The next step came with the installation of electrostatic precipitators. Precipitators were generally designed and predicated on factors evolved from eastern higher sulfur fuels and, all too often, purchased on price only. As a result of these earlier efforts in electrostatic precipitators, many units were found to be entirely too small, reaching, in some instances, only one-half the efficiency that was originally guaranteed by the manufacturer.

Resistivity. Another problem encountered was the complete lack of understanding of the effects of resistivity. Figure 1 is a typical resistivity curve depicting the variance of resistivity with temperature.

To overcome the problem of high resistivity of low sulfur western fuels, two courses were taken. One was to install larger electrostatic precipitators that operate in the traditional 270-300°F range, and the other course was to install hot side electrostatic precipitators generally operating above 700°F. Both approaches have their merits. A well constructed cold side precipitator with sufficient surfaces and low gas velocity can be operated well within design efficiencies. Another approach tried on a very limited basis is to reduce the operating temperature within the electrostatic precipitator to approximately 240°F. The resistivity is reduced at this lower temperature, however you now encounter a very serious dew point problem. Particulates in combination with SO_2 tend to raise the dew point temperature and can cause serious corrosion problems within the ductwork.

On the other hand, the hot gas precipitators attempted to overcome the effect of resistivity by operating at a temperature at which the sulfur content no longer affected the resistivity. It was thought that if one could overcome the resistivity difficulties, a precipitator could then be designed to perform at near clear-stack status. The concept of the hot gas precipitator was encouraging even though this device encountered some difficulties, particularly in the physical design of the structure. At these elevated temperatures, there was, obviously, greater movement in the structural members of the precipitator thus causing warpage and cracking of the precipitator boxes. Although this problem was serious, with re-evaluation of the structural design and greater emphasis placed on the thermal expansion, this problem can and has been overcome. There are some indications that a hot side gas precipitator, under some conditions, seems to develop a coating on the wire emitters which reduces the current emitted and, in some cases, enhances back corona. Additional work needs to be done in this area to determine the extent of this phenomenon.

Gas Conditioning Agents. Gas conditioning agents have proven to be effective in improving collection efficiencies of electrostatic precipitators. Gas conditioning can be applied only with the cold precipitator (270-300°F), since the gas conditioning agent has an effect only on the surface resistivity of the dust particle. The basic

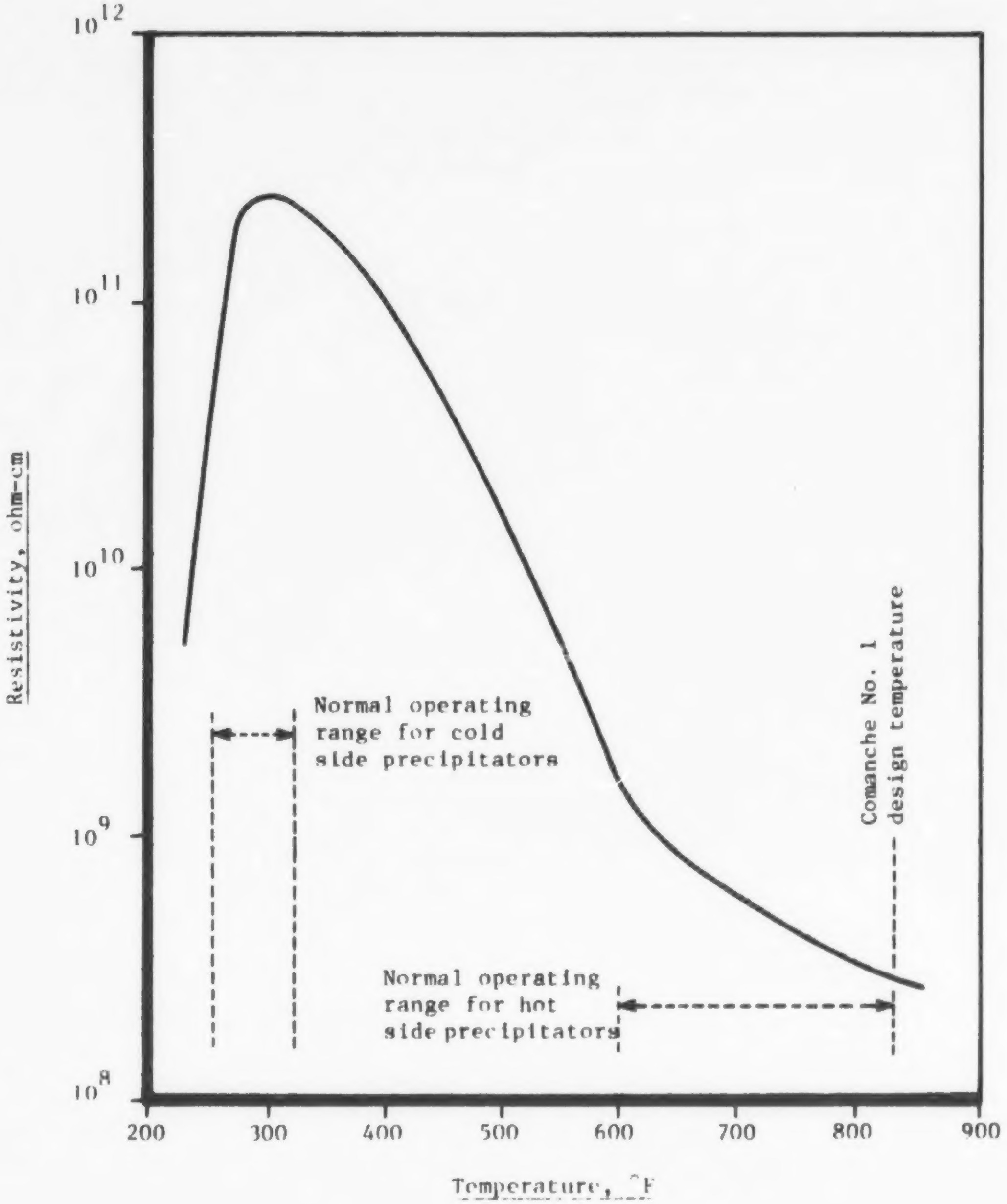

Figure 1. ESP Resistivity Effects

concept of gas conditioning is to inject within the flue gas stream a material such as liquid SO_3 or vaporized sulfuric acid to artificially modify or change the characteristics of the ash. This has the result of allowing the electrostatic precipitator to act as though it were really cleaning an ash from a higher sulfur eastern fuel. Therefore, if the design criteria of the electrostatic precipitator were correctly based on an eastern or higher sulfur fuel, then the limit attained by the addition of the gas conditioning agent will be to approach the limit of operation of the precipitator utilizing the aforementioned fuel.

Gas conditioning has aided in bringing many electrostatic precipitators into compliance and can be an effective method for improving precipitator operation. Gas conditioning was first attempted on a large scale with western coals in 1970 and found to have a relatively good success. If the intent is to convert an existing unit to burn a low sulfur western sub-bituminous coal, gas conditioning can be considered as an aid to an existing precipitator provided that the precipitator is at least large enough to provide marginal efficiency.

In designing new plants, the option between the use of a cold precipitator verus a hot precipitator is up to the individual design considerations of the company and the specific fuel to be used. If it is planned to utilize a large cold precipitator on a new installation, it is highly recommended that the initial design and configuration be predicated on meeting the desired efficiencies without the use of a gas conditioning agent. I strongly urge that gas conditioning be considered only as an aid in helping improve the efficiency of the precipitator if the actual operating efficiency is below the design specifications.

Fabric Filter Baghouse. The fabric filter baghouse can be used for the collection of particulate matter as an alternate to a precipitator. If one forgets the initial baghouse installation at the Southern California Edison station and considers the more recent baghouse installations at Nucla (Colorado) Station of Colorado-Ute and the Sunberry Station of Pennsylvania Power and Light, the success of these recent baghouse operations is attractive. Extensive tests performed at the Nucla Station show overall collection efficiencies greater than 99.9 percent with exit grain loadings of less than 0.005 gr/ft^3.

The real proof of operation of the baghouse is the visual inspections of the stack, and these have been basically clear. The Electric Power Research Institute in conjunction with Meteorology Research Institute and Stearns Roger Inc. recently completed an in-depth study of the Nucla baghouse installation and will publish their results in the near future. I believe it is sufficient to state at this time that the preliminary collection efficiencies of the baghouse exceed the collection efficiencies of the electrostatic precipitator; and I can personally say from observation on my part that during full operation the three stacks of the Nucla Station were absolutely clear. I am not suggesting that a baghouse or any other device is the total answer for particulate control. I do, however, suggest that serious consideration should be given to the fabric filter collection devices not only for new installations but also for retrofitting older units, particularly smaller size units.

Scrubber. The final option to be considered for particulate collection of low sulfur fuels is the use of a wet particulate scrubber. My own company, Public Service Company of Colorado, has five full size particulate scrubbers in operation on units that range from 100 megawatts up to 350 megawatts. In addition to Public Service Company of Colorado, particulate wet scrubbers have been installed at Dave Johnston unit number 4 of Pacific Power and Light, at Clay Boswell and Aurora Stations of Minnesota Power and Light Company, the Four Corners Station of Arizona Public Service Company, and the Holtwood Station of Pennsylvania Power and Light. Table 3 is a list of some of the design parameters associated with these scrubbers. The aforementioned scrubber installations were designed and installed specifically as particulate scrubbers. This is not intended as a complete list of scrubbers that have been installed in the United States during the last four or five years. A more complete list of all scrubbers in operation in the United States is provided by Pedco under contract with the Environmental Protection Agency. Figures 2 through 7 are included to depict flow diagrams of the various particulate scrubbers mentioned previously.

All of the scrubbers at the Public Service Company of Colorado are TCA scrubbers. You will note the difference between Figure 2, Cherokee Station, and Figure 3, Valmont Station, is the attempt to operate one module of the Valmont Station as a SO_2 scrubber. The Valmont scrubber was our first particulate scrubber and was constructed in two separate modules. After several years of operation and facing a more stringent SO_2 standard, one module was converted to a SO_2 mode for experimental purposes. Several things have been determined in this short test period, none of which are particularly surprising. For example, the scaling problem was intensified when the scrubber liquid pH was controlled. Secondly, the maintenance and operation costs were extensively increased in the SO_2 mode. And, finally the question of how to dispose of the tons of sludge produced by the addition of the limestone to the scrubber slurry arose.

From a purely operation and maintenance cost standpoint, the operation of a particulate scrubber far exceeds the cost of operating a precipitator, particularly

Green, 1975.

TABLE 3
SUMMARY OF OPERATING PARTICULATE SCRUBBERS

	Public Service Company of Colorado			Pacific Power And Light	Minnesota Power And Light		Arizona Public Service	Pennsylvania Power And Light
	Arapahoe Station	Cherokee Station	Valmont Station	Dave Johnston Station	Clay Boswell Station	Aurora Station	Four Corners Station	Holtwood Station
No. of Equipped Boilers	1	3	1	1	1	2	3	1
Scrubber Capacity, MW	100	600	100	330	350	116	575	75
No. of Modules	2	9	2	3	1	2	6	1
Vendor	UOP	UOP	UOP	CHEMICO	KREBS	KREBS	CHEMICO	CHEMICO
Type	3 stage TCA	3 stage TCA	3 stage TCA	Venturi	High Pressure Spray	High Pressure Spray	Venturi	Venturi
Start Up Date	11/73	Unit 1 9/73 Unit 3 10/72 Unit 4 9/74	11/71	4/72	5/73	6/71	12/71	5/70
Availability, PCT.	86	Unit 1 55 Unit 3 95 Unit 4 82	76	NA	NA	NA	80	90-95
Particulate Removal, PCT.	96	97	96	99	99	98	99.2	99
SO_2 Removal, PCT.	45	20	45	40	20	20	30	20
L/G, gal/1000 acf	54.2	55.9	58.3	13	8	8	9	12.5
P, in. H_2O	10-18	10-18	10-18	15	4	4	28	5-6
Treated Gas Flow	299,000 SCFM	Unit 1 302,000 SCFM Unit 3 360,000 SCFM Unit 4 897,000 SCFM	299,000 SCFM	1,500,000 ACFM @ 270°F	1,300,000 ACFM @ 254°F	291,160 ACFM @ 340°F ea. boiler	Unit 1 814,000 ACFM @ 340°F Unit 2 814,000 ACFM @ 340°F Unit 3 1,030,000 ACFM @ 340°F boiler	209,000 SCFM
Turndown Ratio, PCT.	47-105	47-105	47-105	30	0-110	0-110	50	NA

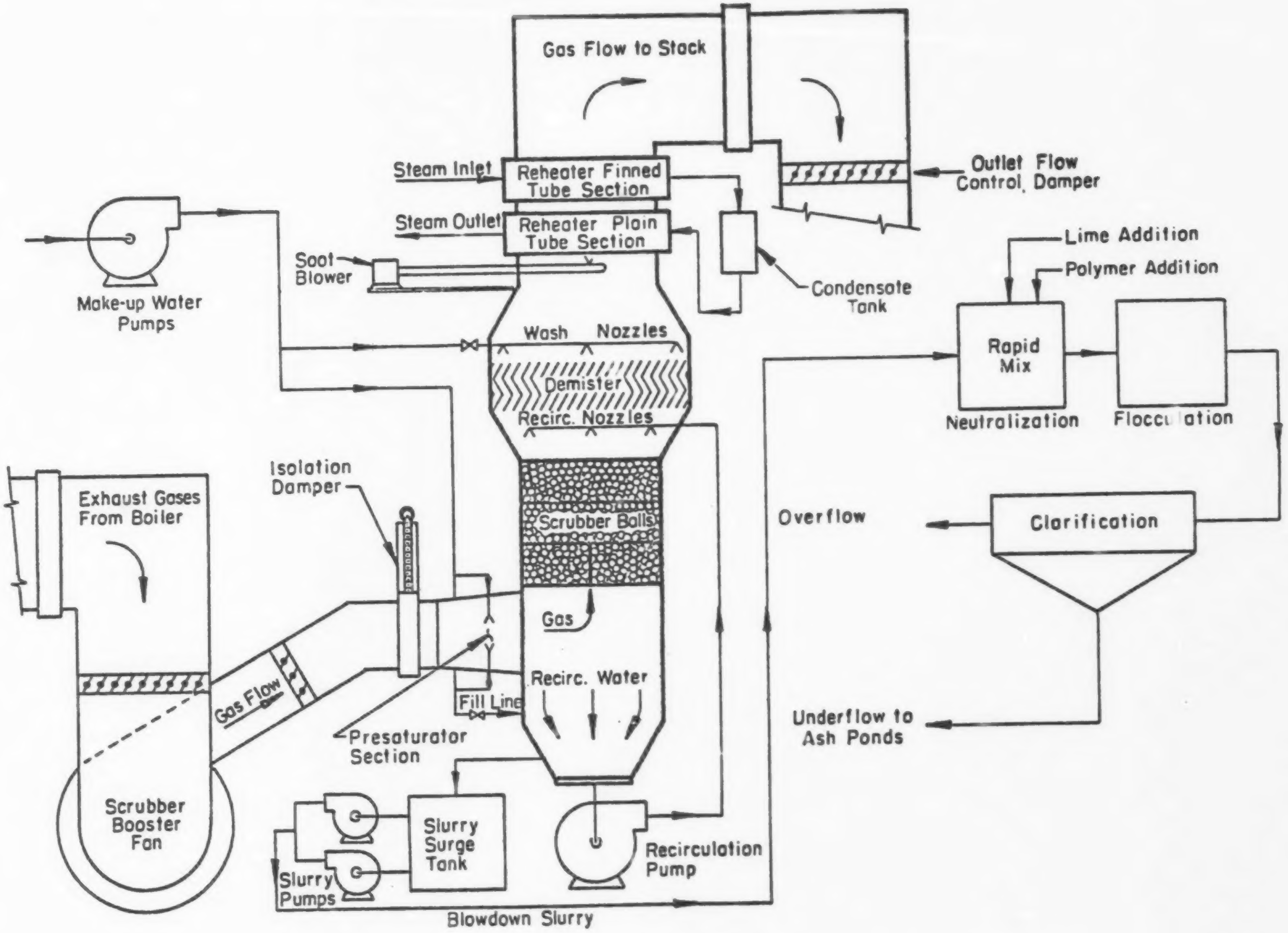

Figure 2. Simplified Cherokee Station Scrubber Flow Diagram

Green, 1975.

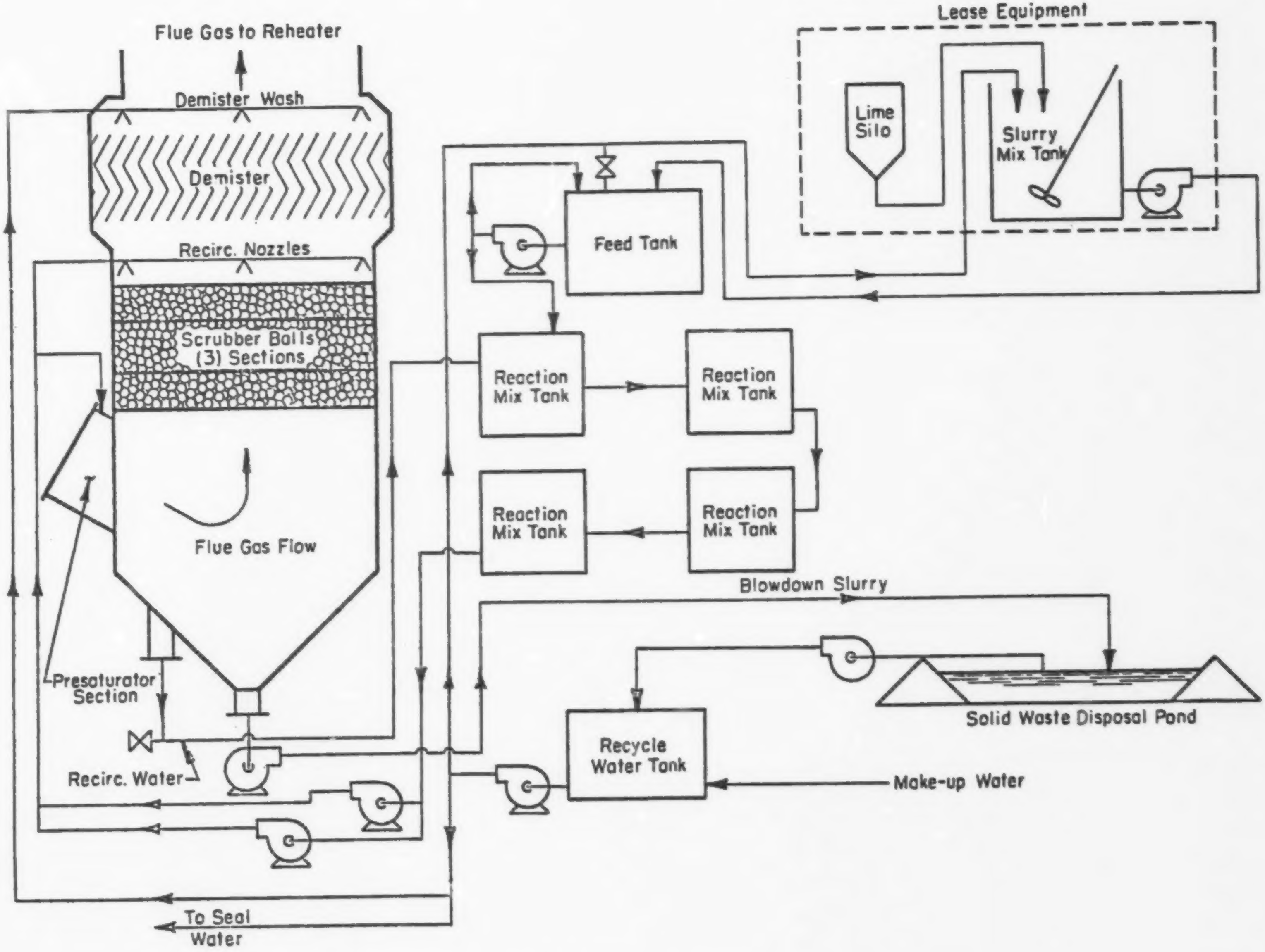

Figure 3 Valmont Station Modified Scrubber Flow Diagram

Green, 1975.

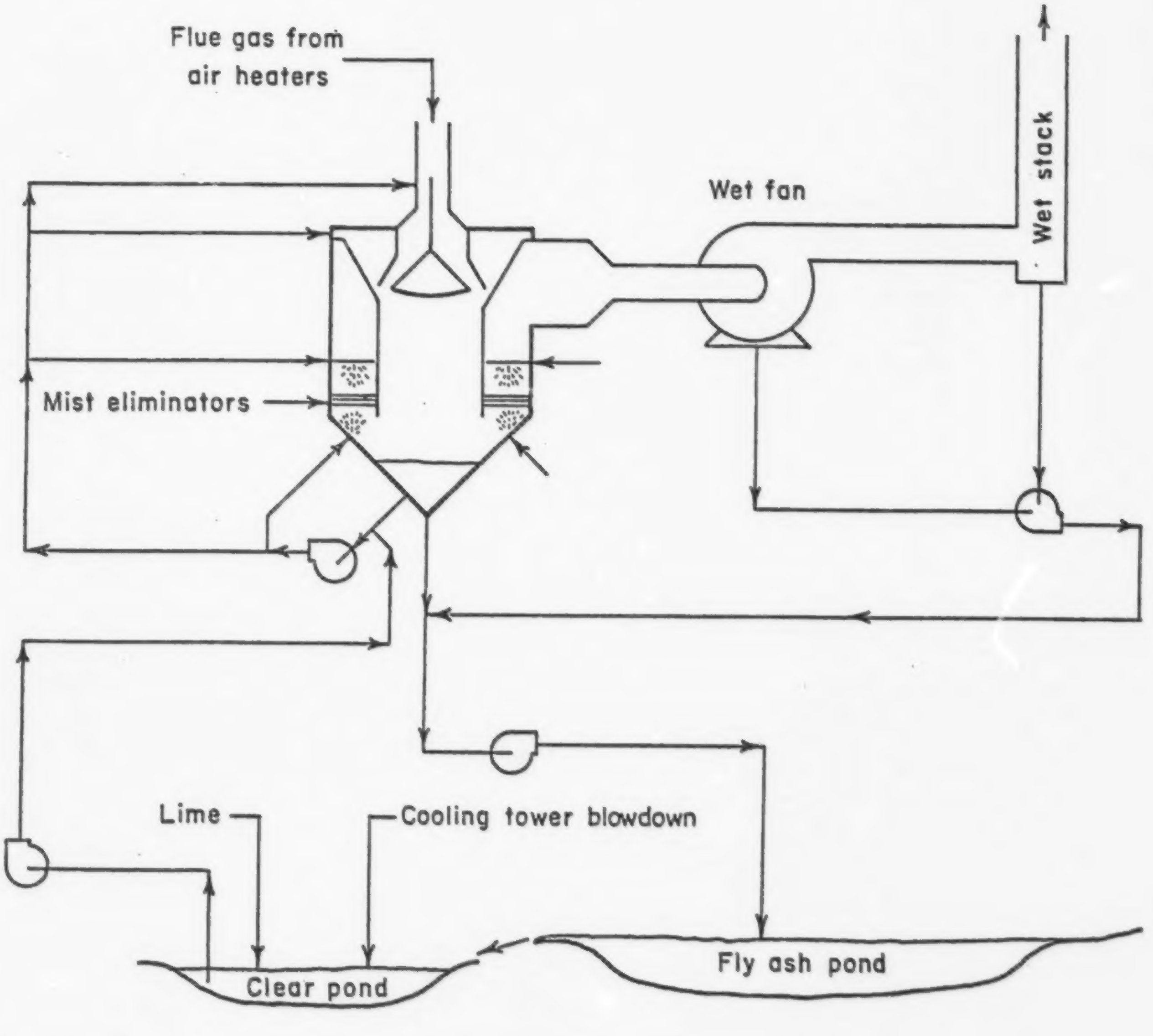

Figure 4 — Simplified flow diagram for the Dave Johnston fly ash scrubbers.

Green, 1975.

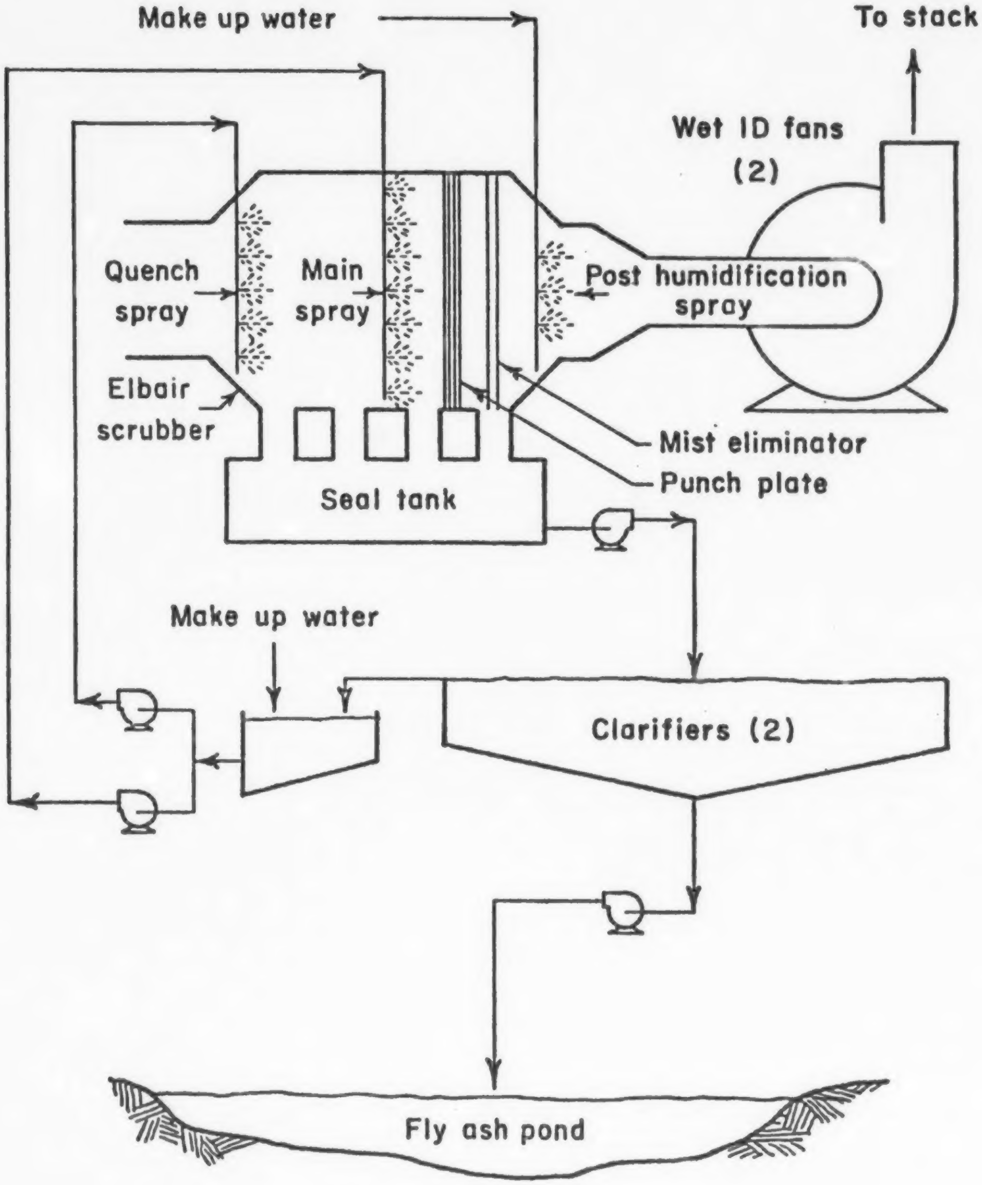

Figure 5. - Simplified flow diagram for the particulate scrubber at the Clay Boswell station.

Green, 1975.

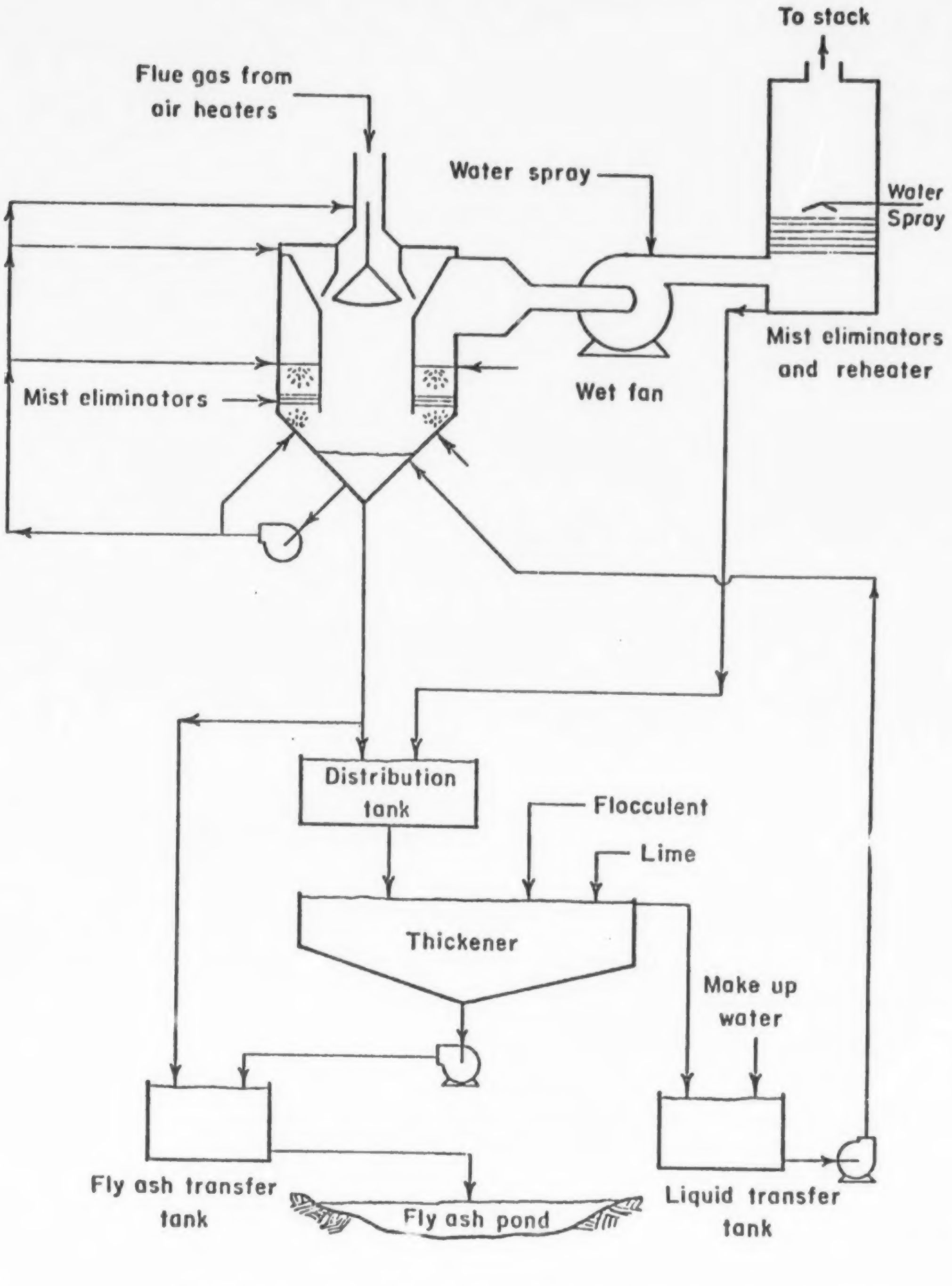

Figure 6. Simplified Flow Diagram for the Four Corners Fly Ash Scrubbers.

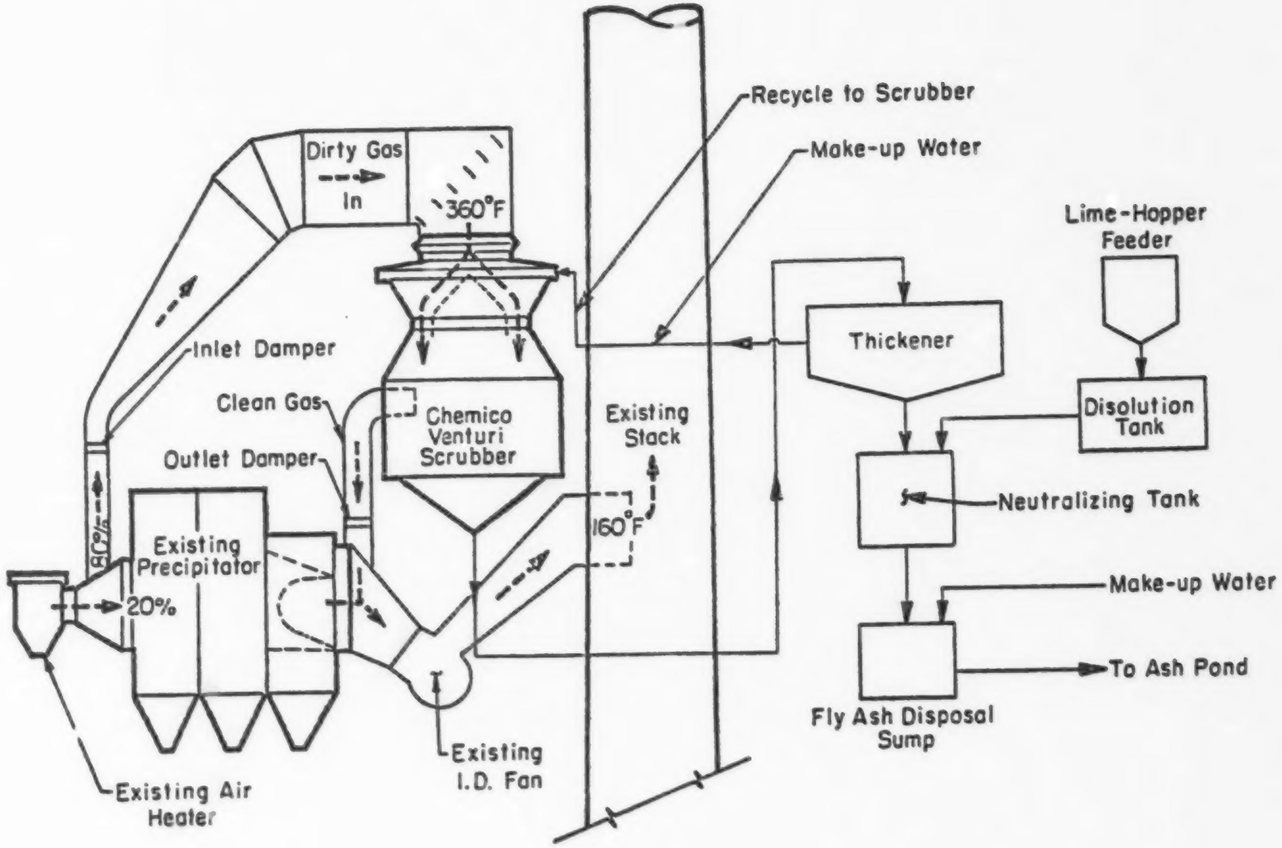

Figure 7. Holtwood Station Scrubbing System General Arrangement

Green, 1975.

in terms of increased maintenance. The particulate scrubbers have proven to be efficient devices in collecting fly ash. They are not subject to resistivity difficulties, wire breakage, plate warpage and many of the other concerns of precipitators. On the other hand, particulate scrubbers introduce a whole host of problems never conceived when operating precipitators. I will not dwell on specific problems associated with scrubbers as there have been numerous papers and discussions on the merits of various scrubbers. I believe it is sufficient to note that when considering the scrubber as a control device, the problems associated with flue gas desulfurization scrubbers are also basically inherent in the particulate scrubber.

CONTROL OPTION CONCLUSIONS

The use of the wet scrubbers for control of particulate matter has proven to be successful, particularly in cases where a low sulfur coal is used as the primary boiler fuel; however, these wet scrubbers have required numerous modifications to improve on-line availability and also require a high degree of maintenance as compared to other particulate control devices. The wet scrubber has also introduced the added dimension of water pollution control for the scrubber effluent.

The operation and maintenance cost associated with the wet scrubber is considerably higher than an electrostatic precipitator to achieve the same collection efficiency. It is for this reason that most utility companies place less consideration on the use of the wet scrubber for particulate control. Many new generating installations will utilize electrostatic precipitators for particulate control, and consider using scrubbers only for meeting future SO_2 emission requirements. It is also surprising to note the renewed interest in fabric filters for particulate control.

New electrostatic precipitator designs for both hot and cold gas precipitators are now proven to be more successful in collecting high resistivity fly ash than were earlier designs. It is my belief that for the collection of particulate matter a precipitator may be the preferred device, particularly when considering operational cost, water pollution and availability problems associated with the wet scrubbers.

During the past twenty years we have seen tremendous improvements in the state of the technology. We have by no means found all the answers to the many problems associated with the control of particulates. Future projects planned by such organizations as the Electric Power Research Institute and many individual utility companies will add further insight into the problem of particulate collection, and, hopefully, we will continue to improve upon these devices. It seems that in this day and age the most important decision to be made in the construction of any new power plant is the determination of the air pollution control equipment to be used. I believe that when burning the lower sulfur western fuels, the decision of the proper control equipment becomes even more difficult to resolve.

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WESTERN COAL USE IN INDUSTRIAL BOILERS

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WESTERN COAL USE IN INDUSTRIAL BOILERS

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ABSTRACT

Six small and intermediate-sized (10,000 to 250,000 lb/hr steam) coal-fired boilers in the upper Midwest have each been tested on both a bituminous eastern coal and a subbituminous western coal.

The purpose of this study was to determine the feasibility of substituting western subbituminous coal for eastern bituminous coal as a means of reducing the SOx emissions from this class of boiler and to demonstrate the feasibility of greatly expanded western coal utilization as a means of reducing the use of oil and gas.

The scope of this study was such that the representative boiler types were tested on both eastern and western coal for a period of time sufficient to completely characterize their individual emission and operational characteristics.

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1.0 INTRODUCTION

Faced with the problem of complying with sulfur dioxide control regulations, electric utilities and industries in the Midwest have been increasing their use of low-sulfur western coal. The extent to which Midwestern demand for western coal will continue to increase depends on a number of factors. Foremost among these are: (1) the evolution of federal, state, and local sulfur dioxide control regulations, (2) the growth of coal as a boiler fuel, and (3) the cost of western coal relative to the costs of alternate fuels and control technologies.

The upper Midwest region (Minnesota, Wisconsin, Iowa, Nebraska, and Illinois) is presently the only area where low-sulfur western subbituminous coal is cost competitive with midwestern and eastern coals. Within this region, there is considerable variation with regard to western coal use versus the traditional eastern supply. This variability is due in part to equipment limitations which dictate that a certain coal be burned.

For this reason it is necessary to determine the operational compatibility of western coal with existing industrial coal-fired equipment, if fuel substitution is to be considered a viable sulfur oxides control strategy.

The purpose of this program, the test results of which are detailed in this paper, was to assess the effectiveness of the use of lower sulfur western coals as a means of reducing sulfur oxides emissions from industrial-sized boilers in the size range 10,000 to 250,000 pounds of steam per hour. The impact on SOx, NOx, CO, particulates, and unburned hydrocarbons emissions has been assessed as a consequence of this fuel conversion.

The scope of the testing program included testing six representative types of coal-fired industrial boilers for a period of one month each on eastern and western coal. During this testing period, the pollutant emissions listed above were measured both in a baseline configuration and in an optimized firing mode. Operational problems of the unit were characterized for each coal. Potential reductions of pollutant emissions have been estimated for each unit type and each coal tested.

2.0 PROPERTIES OF WESTERN SUBBITUMINOUS COALS

A large supply of low sulfur, subbituminous coal exists in the Powder River region of Wyoming and the Fort Union region of southeast Montana. This coal is being mined at a rapidly increasing rate. One mine in Wyoming, for example, increased production from 0.89 million tons per year in 1973 to 3.3 million tons per year in 1974, a factor of 3.7 in only one year. However, the most impressive statistics are the reserve capacity of these western coal fields. That same mine in Wyoming whose production increased so dramatically in 1974 has a reserve capacity of 18.5 billion tons. This translates to a lifetime of 50 years at current production rates. The large reserves, coupled with the relative ease of strip mining, point to a ready supply of coal for fuel if other constraints are met. One of these constraints is the subject of this paper.

The compatibility of these western subbituminous coals with existing industrial boilers could be a hinderance to their wide acceptance as a boiler fuel. The compatibility of coal and boiler are determined both by coal properties and by boiler design. Since the boiler designs are fixed in existing units, the coal properties are the variables of interest.

Western coal characteristics are those of a typical subbituminous coal: an ash-free higher heating value of 8,200 to 10,500 moist Btu/lb, and a high moisture content of 20% to 30%. The ash content of most of these coals is less than 10% by weight. The western subbituminous coals exhibit high volatile to fixed carbon ratios, typically approaching a value of one. A typical, as-received, analysis of a Montana subbituminous coal is given in Table I.

TABLE I

TYPICAL WESTERN COAL CHARACTERISTICS

(Colstrip, Montana)

PROXIMATE ANALYSIS

	As Received	Dry Basis
% Moisture	24.36	xxxxx
% Ash	8.86	11.71
% Volatile	32.53	43.00
% Fixed Carbon	34.25	45.29
	100.00	100.00

ULTIMATE ANALYSIS

	% Weight As Received	Dry Basis
Moisture	24.36	xxxxx
Carbon	51.19	67.67
Hydrogen	3.36	4.44
Nitrogen	0.77	1.02
Chlorine	0.01	0.01
Sulfur	0.88	1.17
Ash	8.86	11.71
Oxygen (diff)	10.57	13.98
	100.00	100.00

FUSION TEMPERATURE OF ASH

	Reducing
Initial Deformation	2130°F
Softening (H = W)	2180°F
Softening (H = 1/2 W)	2205°F
Fluid	2240°F

H is Cone Height
W is Cone Width

SILICA VALUE = 0.09

T250 = 62.42°F

ESTIMATED VISCOSITY
at Critical Viscosity
Temperature of = 2380°F

MINERAL ANALYSIS

	% Weight Ignited Basis
Phos. pentoxide, P_2O_3	0.20
Silica, SiO_2	40.85
Ferric oxide, Fe_2O_3	9.33
Alumina, Al_2O_3	17.36
Titania, TiO_2	0.83
Lime, CaO	11.50
Magnesia, MgO	3.76
Sulfur trioxide, SO_3	15.13
Potassium oxide, K_2O	0.48
Sodium oxide, Na_2O	0.42
Undetermined	0.14
	100.00

The western subbituminous coals are also classed as "free-burning" coals. In the free-burning coals, the pieces do not fuse together, but burn separately or, after fusion, the mass breaks up quickly into fragments. This characteristic causes problems in certain types of stokers where there is inadequate control of the undergrate air distribution. Some specific western coal problems for five types of combustion devices, and the property that causes them, are presented in Table II.

The high moisture content of the western coals causes the greatest combustion difficulty in industrial-sized equipment. In most units with superheaters, it leads to high steam superheat temperatures. It also causes flame stability problems in pulverized coal combustion and ignition problems in stoker-fired units. It is clear that in order to recover the lost steam capacity, some pre-drying of western coals will be necessary, for firing in units designed for eastern coal.

The second major problem with western coal is the size distribution of the delivered coal. Most western coals do not travel or weather well. The coal has a tendency to break into fine sizes while in transit. Therefore, even if the coal has been sized before shipment, the as-received coal will exhibit a shift in size distribution toward the smaller sizes. This shift becomes more severe with longer transit and/or storage periods. The effect of this coal property on stoker unit performance is discussed below.

This paper is divided into a discussion of pulverized coal firing and stoker firing of both eastern and western coal. A general overview of boiler performance is presented in Table III. Here, the units tested are rated in terms of emissions, efficiency, and overall ease of operation. Comments are presented where appropriate. The type and source of the coals tested are also given for each boiler.

3.0 PULVERIZED COAL COMBUSTION

The pulverized coal-fired boiler tested was Unit No. 3 at Dairyland Power Cooperative at Alma, Wisconsin generating station. This four burner face-fired unit manufactured by Riley Stoker Corporation is rated

TABLE II

SOME SPECIFIC WESTERN COAL PROBLEMS

		COMBUSTION DEVICES							
	PROPERTIES	VIBRATING GRATE STOKER	SPREADER STOKER	PULVERIZER	UNDERFED STOKER	OVERFED STOKER	TRANSPORTATION	STORAGE	ASH COLL/ DISPOSAL
COAL	Low Heating Value		o Flame stability o Reduced capacity	o Reduced capacity o High maintenance	o Reduced capacity	o Reduced capacity	o Lower Btu's/ ton-mile	o Increased equipment loading	o More ash/Btu
	High Moisture	o Ignition problems	o High superheat temperature	o Flame stability o High superheat temperature o Poor grinding	o Poor ignition o Nonuniform bed thickness	o Poor ignition	o Freezing	o Fires	o More ash/Btu
	Low Sulfur								o Inefficient precipitators
	Dustiness/ Fineness	o Uneven fuel bed	o Poor grate coverage o Overheating of grates	o Feeder plugging	o Uneven fuel bed with uncovered grate areas	o Carbon carryover o Undergrate air restriction	o Coal loss o Fugitive dust	o Fires o Fugitive dust	
	Friability		o Ash pit fires		o Ash pit fires	o Ash pit fires			
ASH	Free Burning	o Uncovered grate			o Uncovered grate	o Uncovered grate			
	Slaking/ Weathering						o Coal size degradation	o Coal size degradation	
	Na and Ca		o Increased fouling	o Fouling o Sintering					o Caking

TABLE III

DESIGN TYPE OF UNITS TESTED AND OVERALL PERFORMANCE ON EASTERN AND WESTERN COALS

TYPE OF UNIT TESTED	OVERALL PERFORMANCE RATING PER COAL				COMMENTS	EASTERN	WESTERN
	Good	Fair	Poor	Unacceptable			
PULVERIZED COAL (Riley) o 230,000 lb/hr steam o Four Burner Face-Fired o Two Ball Tube Mill Pulverizers o UOP ESP	Eastern and Western				Reduced maximum capacity	Western Kentucky (River King)	Montana Sarpy Creek (Westmoreland)
VIBRATING GRATE STOKER (Detroit) o Water-Cooled Grate o 45,000 lb/hr Steam o FD Fan/Natural ID o Cinder Trap Partic. Removal	Eastern	Western			Improved coal sizing would improve performance	Western Kentucky (Vogue)	Wyoming (Big Horn)
TRAVELING GRATE STOKER (LaClede) o 60,000 lb/hr Steam o FD Fan/Natural ID o No Particulate Controls	Eastern		Western		Severely affected by coal size	Western Kentucky (Vogue)	Wyoming (Big Horn)
UNDERFED STOKER (Westinghouse) o Multiple Retort o 100,000 lb/hr Steam o Cyclone Dust Collector		Eastern		Western	Specially sized western coal was used for the test, however, the unit would not respond to load demand. Modifications are necessary to undergrate air system in order to burn western coal.	Kentucky and Illinois	Wyoming (Big Horn)
SPREADER STOKER (Detroit) o 160,000 lb/hr Steam o Traveling Grate o Multiclone Cyclone o FD and ID Fans o Superheat, Economizer, and Air Heater		Western	Eastern		Maximum load reduced to 130,000 lb/hr steam on western coal due to high superheat temperatures. Large carbon losses on eastern coal--smoking.	Southern Illinois	Montana (Colstrip)
SPREADER STOKER (Westinghouse) o 100,000 lb/hr Steam o Traveling Grate o FD and ID Fans o Superheat, Economizer	Western and Eastern				Able to maintain full load on western coal	Kentucky (Vogue)	Montana (Colstrip)

at 230,000 lb/hr steam flow. The coal is pulverized with two ball tube mills, one mill for the upper two burners and one for the lower two burners. The unit is equipped with a steam spray attemperator. Fly ash collection is accomplished with a UOP designed cold side electrostatic precipitator (ESP).

The two fuels used during the testing were:

Western Kentucky Coal

- 4% sulfur
- 16% ash
- 10,500 Btu/lb
- 18% volatiles

and

Montana Coal

- 0.77% sulfur
- 12% ash
- 8,400 Btu/lb
- 37% volatiles.

3.1 Boiler Performance - Alma Unit No. 3

The boiler performed well on both coals, although the unit was limited in maximum load due to excessive superheat steam temperature. The steam attemperation system was not adequate to reduce the temperature to the desired 900°F level at loads above 174,000 lb/hr steam on western coal. This compares to a maximum load of 204,000 lb/hr steam on eastern coal. The boiler is design rated at 230,000 lb/hr steam, however this load is no longer achieved.

The primary factor causing the excessive steam temperature is the high moisture content of the coal. This water reduces the flame temperature which in turn reduces the radiant heat flux to the water walls resulting in lower steam generation. This lower heat transfer (a function of temperature to the fourth power) removes less heat in the radiant section, however, the gas still contains a large enthalpy which then acts on a decreased amount of steam in the convective section resulting in increased steam temperatures. The water in the fuel also results in greater gas flows which increase heat transfer rates in the convective pass.

The excessive steam temperature problem is a function of boiler design. For example, a boiler designed for western coal might not be able to make design steam temperature on eastern coal.

Increased steam attemperation would result in full capacity operation on western coal.

3.2 Pulverizing Mill Performance. Eastern coking coals, when exposed to furnace temperatures, will swell and form lightweight, porous coke particles. These may float out of the furnace before they are completely burned. As a result, carbon loss will be high unless pulverization is very fine. Free-burning (western) coals, on the other hand, do not require the same degree of fineness because the swelling characteristic is absent.

High-volatile (western) coals ignite more readily than those with a low volatile content. Therefore, they do not require the same degree of fine pulverization. With the exception of anthracite, however, the low-volatile coals are softer, and may be said to have a higher grindability. As a result, mill capacity is greater at increased fineness than with high-volatile coals.(1)

Table IV shows the screen analyses and the loads of the coal burned in tests 9, 16, 57, 63, 75, and 78. Tests 9 and 16 were on eastern coal. Test 16 was with one mill, a total of two, operating so the load in the mill was the same as it would have been with both mills operating at 104,000 lb/hr steam. The screen analyses of tests 16 and 78 may then be compared. It is seen that the western coal did not grind quite as well as the eastern coal. However in the opinion of Reference 1, free-burning coals need not be ground as fine as coking coals, and this was not thought to be a severe problem. An equally important factor in mill grinding capacity is moisture. From Reference 1, frequently too much emphasis is placed on grindability, while other factors such as moisture, which also affects mill capacity, are almost entirely overlooked. The capacity of a pulverizer is not directly proportional to the grindability of a coal. Correction must be made for variation in fineness and moisture content.

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1. Combustion Engineering, O. de Lorenzi, editor, Combustion Engineering Co., Inc. 1947, p. 7, 8.

TABLE IV

SCREEN ANALYSES OF PULVERIZED COAL

	EASTERN		WESTERN				
Test No.	9	16	57	63	63	75	78
Load, thousands lb/hr steam	125	52	170	131	94	160	110
- 80 mesh, %	0.65	1.00	2.90	1.40	0.65	8.75	1.53
- 80 +100 mesh, %	0.75	1.00	2.30	1.30	2.32	7.70	2.23
-100 +140 mesh, %	2.75	3.35	20.65	4.05	8.91	21.65	6.60
-140 +200 mesh, %	6.95	8.20	34.60	8.30	32.26	16.75	17.46
-200 mesh, %	88.90	86.45	39.55	84.95	55.66	45.15	72.18
Moisture, %	5.15	2.20	22.05	12.75	17.94	19.33	17.75

Without quantitative analysis, it can be seen in Table IV that the moisture content of test 78 is a factor of 8 times higher than test 16. The grindability of the other tests tend to follow the moisture content. Test 57 with the highest moisture content exhibited the poorest grindability, followed by tests 75, 73, 78, and 63 in order of increasing grindability.

The poorly pulverized coal burns more slowly resulting in lowered heat transfer in the near-flame region (radiant section) and increased heat transfer to the convective section. At high loads (tests 57 and 75), the poor grind probably contributed to the excessive superheat steam temperature problem.

3.3 Emissions from Alma Unit No. 3

A coal performance comparison for Alma Unit No. 3 is presented in Table V. In this table, western coal, test 66, is compared to the nearly identical eastern coal, test 9.

Significant differences in coal performance are noted for:

- SOx emissions
- NO emissions
- Carbon carryover

TABLE V

COAL PERFORMANCE COMPARISON, ALMA UNIT NO. 3

	Western	Eastern
Test No.	66 ESP Inlet	9 ESP Inlet
Load, Klb steam/hr	130	130
Excess O_2, %	3.4	3.4
SOx at 3% O_2, ppm	996	3283
NO, dry at 3% O_2, ppm	372*	490**
CO, at 3% O_2, ppm	31	21
Particulate, lb/10^6 Btu	6.77	7.84
ESP Efficiency, %	99.6	99.6
Carbon Carryover, % by wt	0.55	4.13
Unburned HC, at 3% O_2, ppm	25	31
Boiler Efficiency, %	85	--

*0.52 lb/10^6 Btu

**0.69 lb/10^6 Btu

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In each case the western coal performed better than the eastern coal. Sulfur oxides emissions were reduced by a factor of 3 by substituting western coal. At the same time nitric oxide emissions were reduced 24% and carbon carryover was virtually eliminated. The electrostatic precipitator's performance was not affected by the fuel switch. It continued to operate at 99+% efficiency. Carbon monoxide and unburned hydrocarbon emissions were generally less than 100 ppm each. In the optimum furnace configuration, these emissions are controlled by excess air. Below three percent (3%) excess O_2 in the flue gas, these emissions became significant. Soot formation, resulting in a black stack, was also a problem below 3% excess O_2.

3.4 Discussion

Figure 1 is a plot of nitric oxide as a function of excess O_2 in the flue for western coal at four loads. Figure 2 contains the same type data for the base eastern coal. Both figures show increasing NO with increasing O_2 at a constant load, however, the absolute magnitude of NO emissions from western coal is less at any given load and O_2. Most of the NO data on Figure 1 fall below the EPA limit for new coal-fired units of 0.7 lb of NOx as NO_2 per million Btu (about 500 ppm). Attempts to reduce the NO emissions of the eastern coal to these same (less than 500 ppm) levels resulted in high CO emissions.

Included in the factors that influence NO emissions are:

- Flame temperature
- Fuel nitrogen
- Excess oxygen

In order to control CO emissions from the eastern coal it was necessary to operate at higher O_2 levels; this led to higher NO emissions. For western coal firing, it has been shown that the furnace can operate at lower excess O_2, thus lower NO. Western coal also contains less bound fuel nitrogen than eastern coal. This fuel nitrogen can be as little as half the amount found in typical eastern coals. Assuming complete conversion of this fuel nitrogen to NO, it can be seen that western coal has a distinct advantage over eastern coal.

The third factor affecting NO emissions is flame temperature. The high moisture content of western coal causes the temperature of the western coal flame to be lower than the eastern coal flame. This lower flame temperature lowers both the fixation of molecular nitrogen in the combustion air as well as reducing the fuel nitrogen conversion to NO.

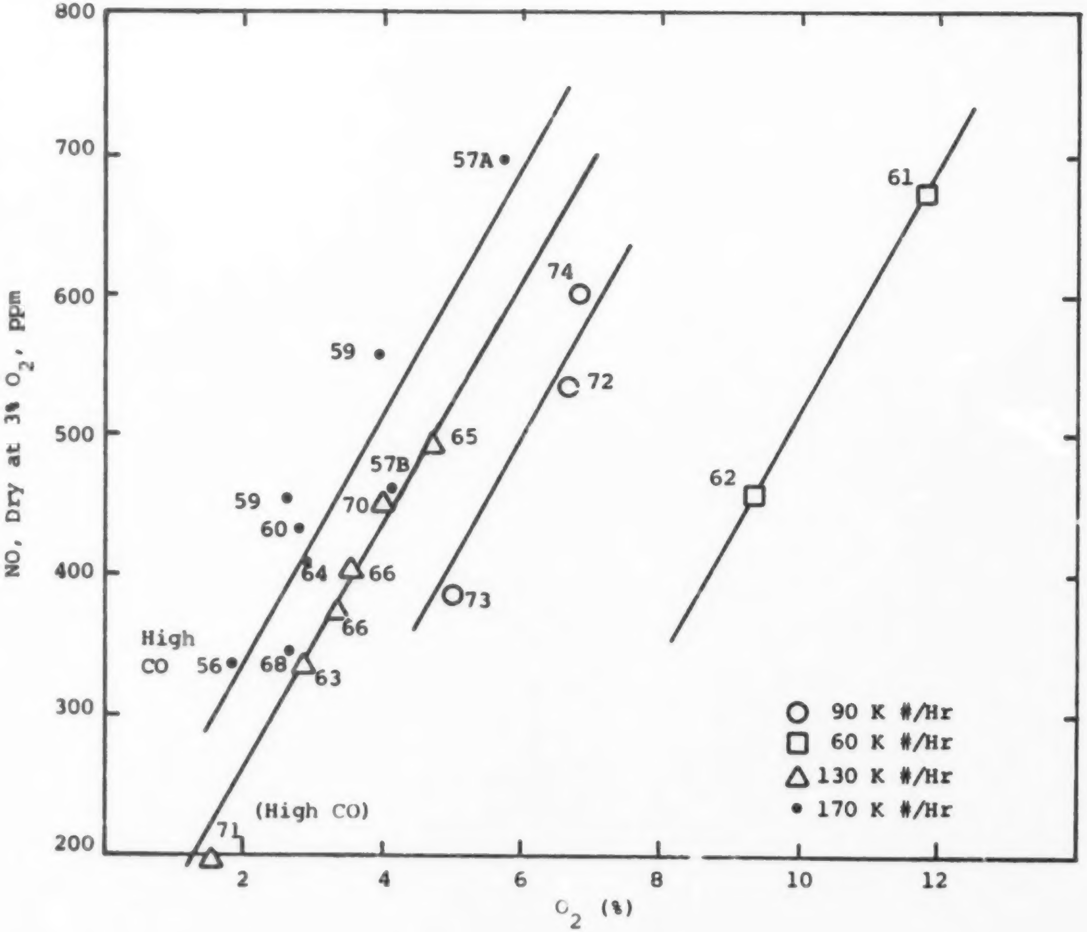

Figure 1. Nitric Oxide vs. Oxygen - Alma #3, Western Coal.

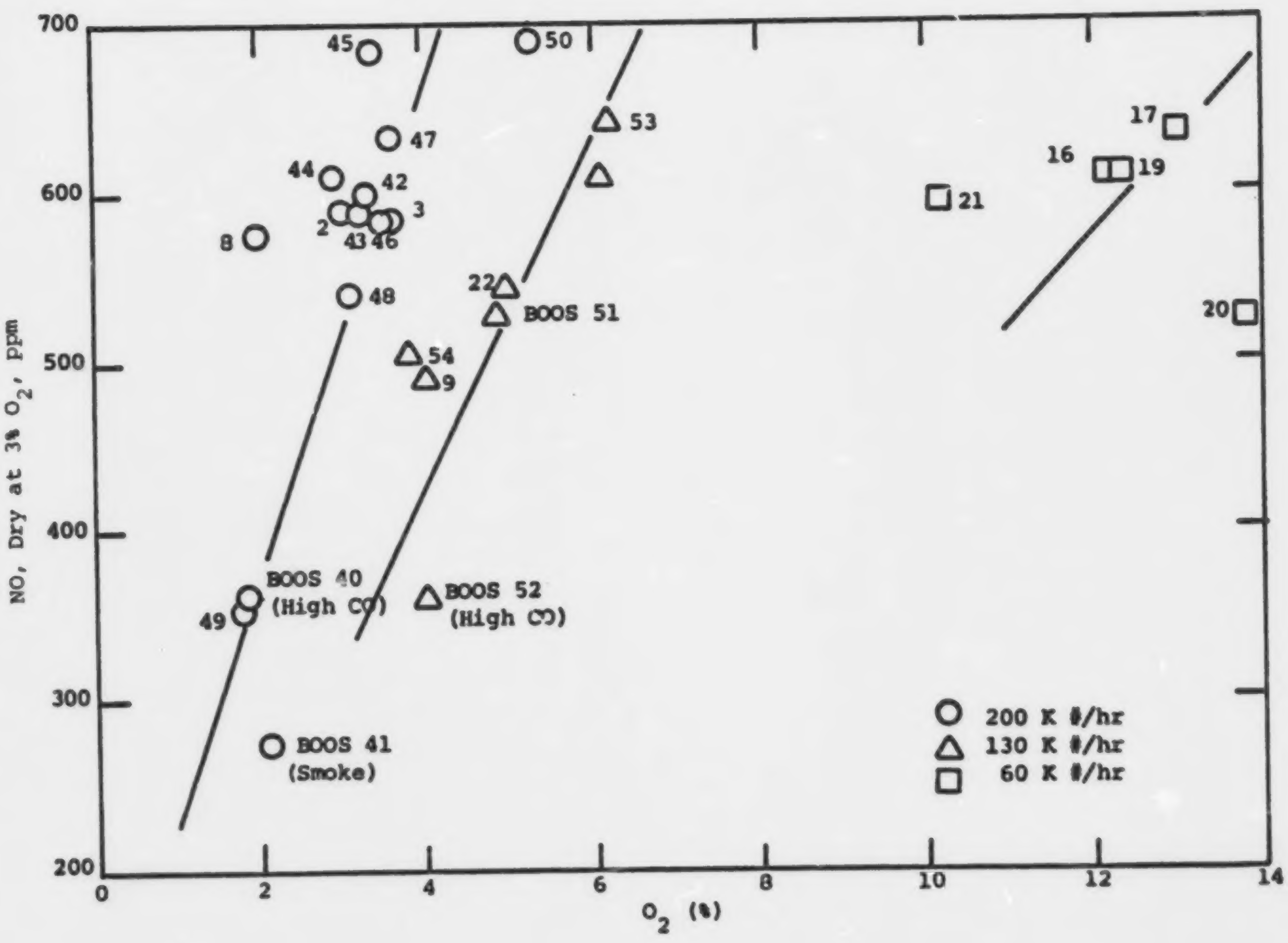

Figure 2. Nitric Oxide vs. Oxygen - Alma #3, Eastern Coal.

Sulfur oxides emissions are largely a function of sulfur in the fuel. There has, however, been some work that indicates that coal ash composition may affect the amount of sulfur oxides emitted (Ref. 2). The comparison of the eastern and western coals on this unit (see Table VI) show the benefit of fuel substitution in the control of SOx emissions.

4.0 STOKER-FIRED BOILERS

Coal firing of industrial boilers can be separated into two broad classes--suspension firing and grate firing.

Suspension firing is normally applied in larger sized units, however, units as small as 35,000 lb/hr steam have been built for pulverized coal firing. Current economics would indicate a break-even point in the 200,000 to 250,000 lb/hr steam flow range. Suspension firing includes both pulverized coal firing (70% through a 200 mesh screen) and cyclone firing (crushed to 1/4" with about 10% through a 200 mesh screen).

Grate firing comprises three general stoker types:

- Underfed
- Overfed
- Spreader

Within these three types, there are a number of variations in feed methods and grate design. Stoker-fired boilers have been built covering the entire capacity range of this study, 10,000 to 250,000 lb/hr steam. The present stoker-fired boiler population represents a highly individualized array of equipment.

Table III, presented previously, lists stoker types tested in this study. From this assortment of units, the emissions and operating characteristics of western coal firing have been determined.

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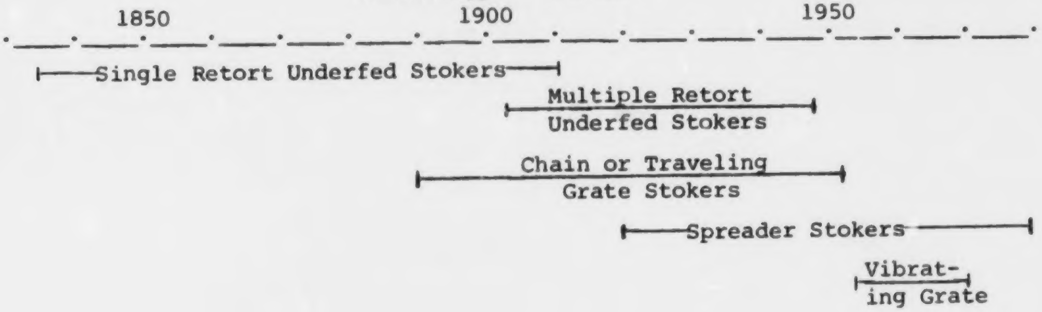
2. Grouhovd, G. H., Tufte, P. H., and Selle, S. J., "Some Studies on Stack Emissions from Lignite-Fired Power Plants," presented at the 1973 Lignite Symposium, Grand Forks, ND, May 9-10, 1973.

4.1 Operational Characteristics of Western Coals in Stokers

A chronological display of the development of the various stoker types would appear as follows:

Chronology of Stokers



Two properties of western subbituminous coals result in operational problems for stoker-fired units. They are:

- Coal weathering - resulting in size reduction
- "Free burning" characteristic - resulting in an uncovered grate

Many older underfed and traveling grate stokers were manufactured with insufficient control of the undergrate air to use western coal as a fuel. The problem manifests itself when a dark spot of unburned coal develops on the grate. This patch of coal can grow into a large clinker if special measures are not taken to remove it. The "black patches" occur because there is insufficient local air pressure under the patch to maintain vigorous burning. The loss of local air pressure occurs because some other portion of the grate, in the same plenum control area, has become thin or bare and allows the combustion air to pass through easily. These units were designed for an eastern coal that formed some coke while burning and in turn maintained even coverage of the grate. The free-burning western coals, on the other hand, tend to form a fine powdery ash which either blows off or falls through the grate, leaving it bare. This problem is compounded by the serious size reduction that occurs while in transit. The small coal particles burn more rapidly when there is available air, however, when there is insufficient air they tend to plug the grate and fuel bed openings and form dark patches which turn into clinkers.

The older underfed and overfed stokers designed for eastern coal will have to have modifications to the undergrate air chamber to allow finer control of the air distribution, if western coals are to be used.

Spreader stokers are affected by the same coal properties but to a lesser extent since approximately half of the combustion takes place in suspension. This suspension burning reduces the number of "fines" that reach the grate. However, the fines in the coal tend to burn close to the spreader, sometimes flashing back into the feeder opening. This flash back mode can be dangerous since there is the possibility of a fire in the coal feeder. Coke and slag also have a tendency to build up on the spill plates and rotor blades if the flash back is allowed to persist. This problem can be alleviated somewhat by rotor speed and spill plate adjustments.

The western coal performed well in the spreader stoker units. In some units designed for eastern coal, the maximum attainable load was about 80% while on western coal. This was due to insufficient induced draft fan capacity and as in pulverized units, high superheat steam temperatures. Removing the major part of the moisture from the coal prior to combustion would alleviate both of these problems.

4.2 Emissions from Stoker-Fired Units

4.2.1 Sulfur Oxides. The emissions of sulfur oxides from stokers is as expected, governed by the sulfur in the fuel. These emissions are independent of load and excess air in the flue gas. Table VI contains the results of a SOx emission comparison for all the coals tested in this study. The overall average SOx emissions from actual operating industrial type boilers decreases from 5.19 $lb/10^6$ Btu on eastern coal to 2.34 $lb/10^6$ Btu on western coal, or 55%. This is to be compared to the reduction as estimated from the fuel analysis from 6.43 $lb/10^6$ Btu on eastern to 1.76 $lb/10^6$ Btu on western coal, or 73%. The sulfur content of the fuel was calculated from published values and not from analyses of actual fuels burned. When the results of the fuel analyses are obtained, the calculation will be corrected to those values.

TABLE VI

SOx EMISSION COMPARISON FOR EASTERN AND WESTERN COALS

WESTERN COAL	Avg Fuel* Sulfur, %	Avg SO_2 Emissions $lb/10^6$ Btu
Montana (Sarpy Creek)	0.80	1.86
Wyoming (Bighorn)	0.60	1.18
Montana (Colstrip)	0.85	1.8
Montana (Colstrip)	0.88	2.56
Wyoming (Bighorn)	0.60	1.28
Overall Average	0.746	2.34
EASTERN COAL		
Kentucky (River King)	3.96	6.46
Kentucky (Vogue, Seam II)	3.22	4.73
So. Illinois	1.9	--
So. Illinois (Stonefort)	3.1	4.92
W. Kentucky (Vogue)	3.22	4.66
Overall Average	3.375	4.19

Average SOx Reduction (Flue Emissions) = 2.85 $lb/10^6$ Btu (55%)

Average SOx Reduction (Est. from Fuel)* = 4.67 $lb/10^6$ Btu (73%)

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*Based on published values, not on actual analyses of fuels used in the tests.

4.2.2 Nitric Oxides. Nitric oxide emissions from stokers exhibited a similar dependence on excess O_2 in the flue gas as the pulverized coal firing. At constant load, nitric oxide emissions increased with increasing excess O_2, as well as increasing slightly with increasing load.

However, the slope of the NO vs. O_2 curve is less for stoker-fired units than for the higher intensity combustion devices. Figure 3 shows an interesting NO vs. O_2 result for a water-cooled vibrograte stoker. The western coal (Wyoming Bighorn) has a slope of 12 (ppm NO/% O_2) compared to the eastern coal (Kentucky Vogue) which has a slope of 35 (ppm NO/% O_2). Figure 4 gives the NO vs. O_2 plot for the same two coals on an overfed traveling grate stoker without a water-cooled grate. On this unit, both coals exhibit the same NO vs. O_2 dependence. In fact, of the boilers tested, the water-cooled grate was the only unit having different NO vs. O_2 slopes for the two coals tested. It is speculative as to whether the additional cooling of the grate affects the conversion of fuel bound nitrogen to NO.

Stokers have overall lower NO emissions than pulverized coal units since the stokers operate in a "staged combustion" configuration. The stokers that have little or no suspension burning such as underfed and overfed stokers have a greater degree of staging than do the spreader stokers. In the stoker, the fuel devolatizes in the fuel bed under reducing conditions, then is mixed with the combustion air above the bed. Mixing is provided by overfire air jets or by front or rear arches in the furnace. Clinkering in the fuel bed establishes a limit to the degree of staging that can be achieved on stokers. Figure 5 shows these limits for a 100,000 lb/hr steam spreader stoker.

Carbon Monoxide and Unburned Hydrocarbons. Carbon monoxide (CO) and unburned hydrocarbons (UHC) emissions from stokers as with all combustion systems, can be controlled by providing adequate excess air and proper mixing to insure complete combustion. High excess air conditions can cause CO and UHC emissions as well as too low excess air. Figure 6 gives the results of CO emission measurements on a 100 Klb/hr steam spreader stoker as a function of excess air for both eastern and western

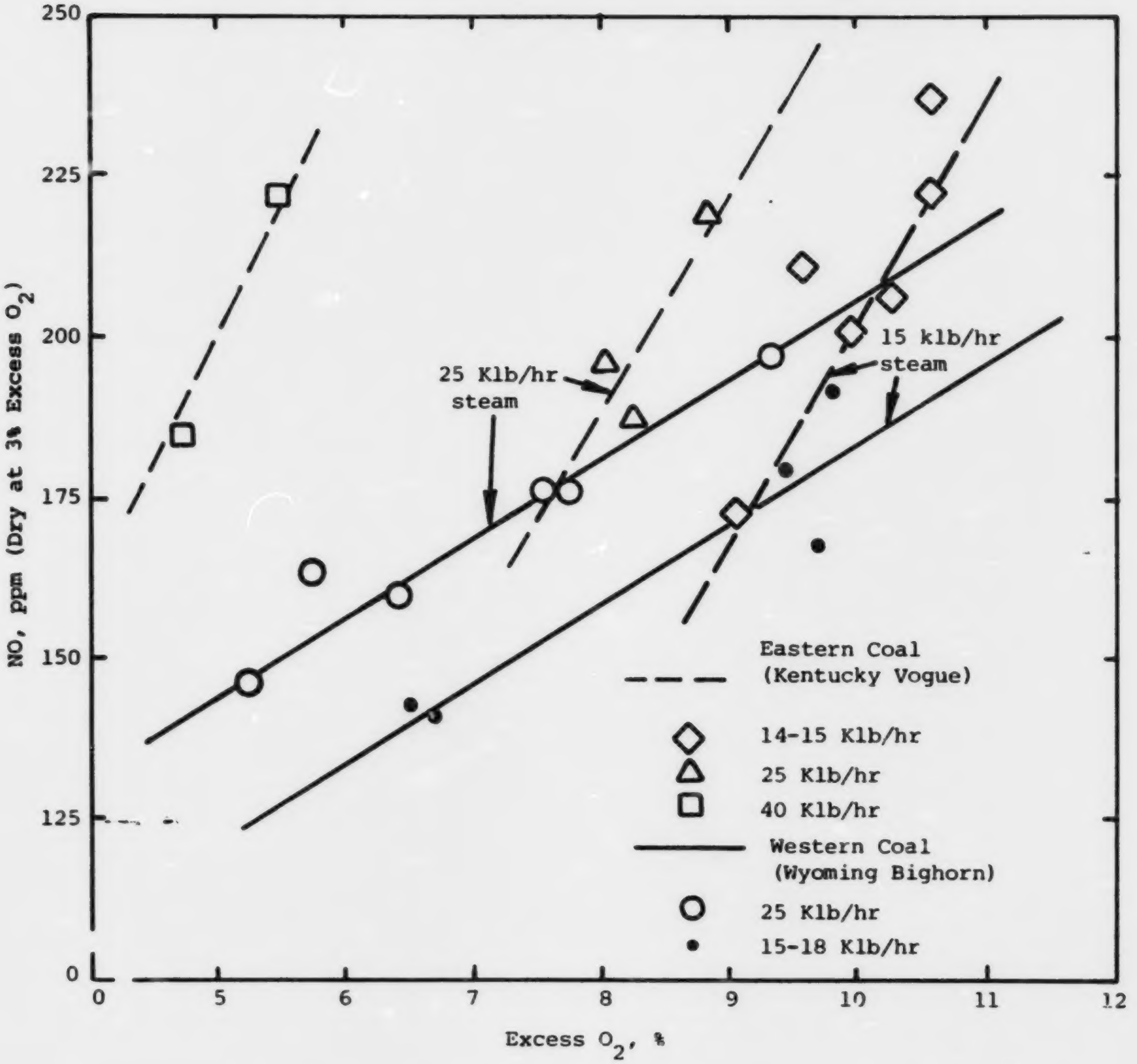

Figure 3. Water-Cooled Vibrograte Stoker, 45,000 lb/hr Steam. Comparison of Western and Eastern Coal Nitric Oxide Emissions (University of Wisconsin-Stout).

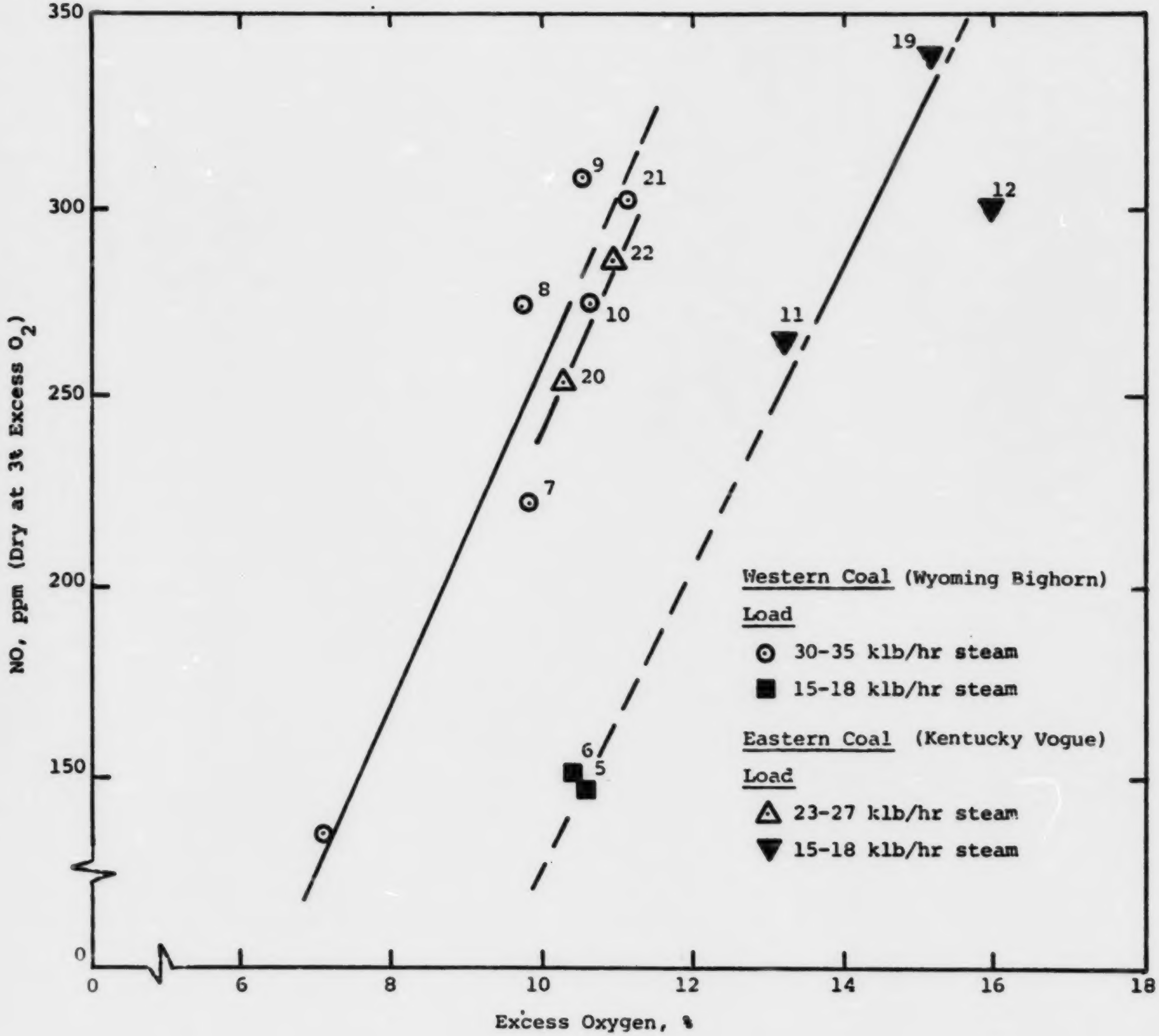

Figure 4. Overfed Traveling Grate Stoker, 60,000 Klb/hr Steam (University of Wisconsin-Eau Claire).

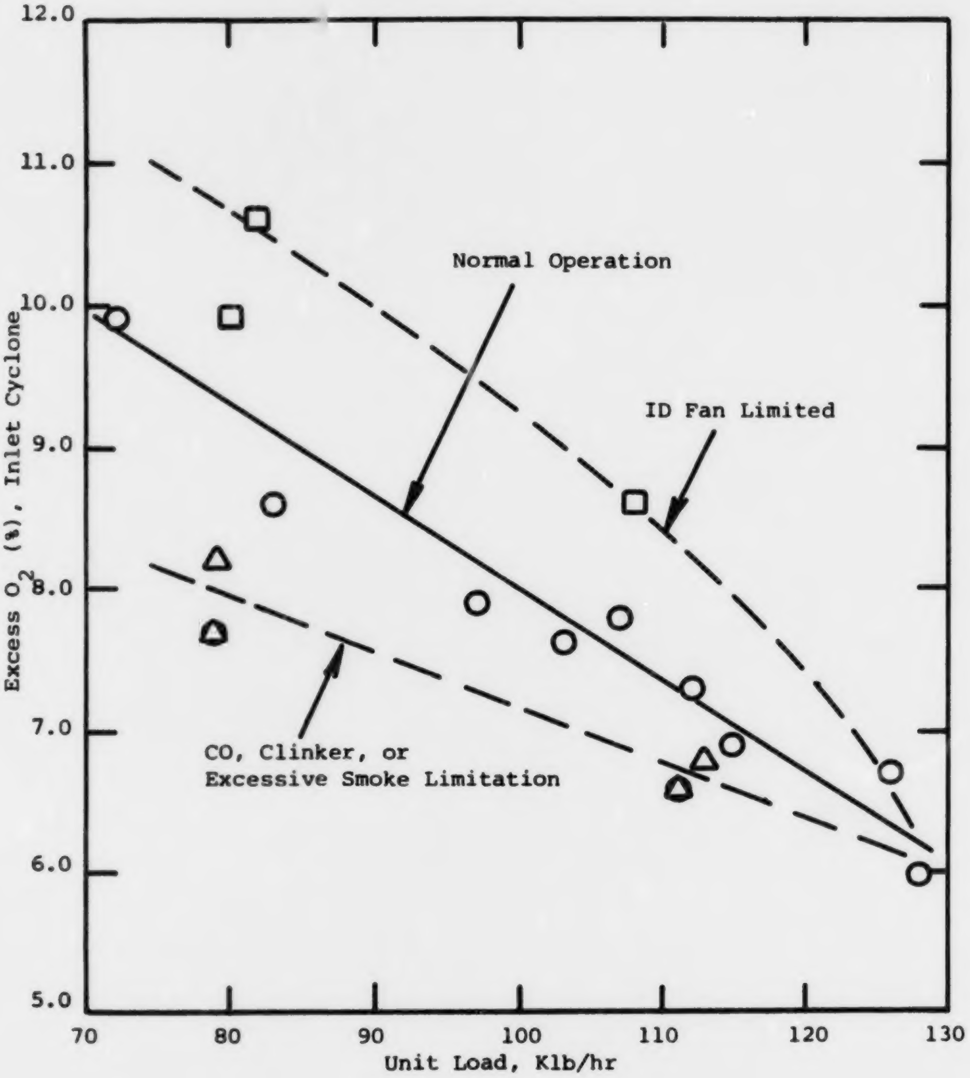

Figure 5. Stoker Firing Staging Limits, 100,000 lb/hr Steam, Western Coal (Willmar Unit 3).

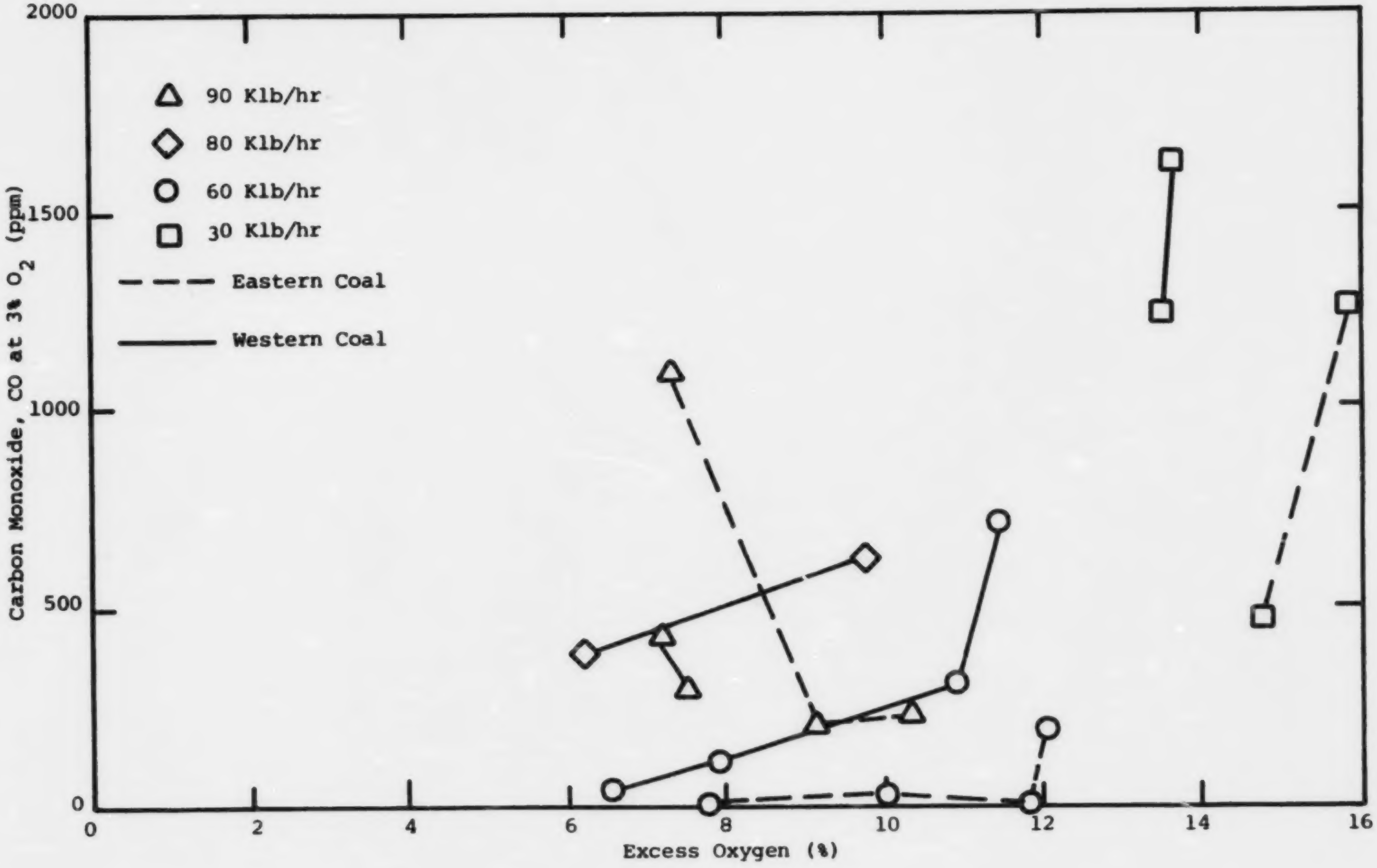

Figure 6. Carbon Monoxide as a Function of Excess Oxygen (University of Wisconsin/Madison).

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coal. At high load (90 Klb/hr steam), CO emissions increase with decreasing excess air; however, at low and intermediate loads, a point is reached where increasing excess air results in rapidly increasing CO emissions. This behavior was observed for both coals. At low excess air, CO results from inadequate mixing of fuel and air. At high excess air settings, the fuel bed is thin even to the extent of some uncovered grate area which is thought to lead to local quenching of the flame by the combustion air and incomplete oxidation of CO to CO_2. The western coal can be fired at 2% lower excess O_2 at high load while producing comparable levels of CO emissions. This means higher unit efficiency because of lower dry gas and combustible losses.

Table VII contains unburned hydrocarbon emission data for both an eastern and a western coal on the same spreader stoker unit described above. Unburned hydrocarbon emissions were higher at low load and high excess air than at high load and normal excess air, thus following the same trends as the CO emissions.

TABLE VII

A COMPARISON OF HC EMISSIONS FROM EASTERN AND WESTERN COALS

Load (klb/hr)	Eastern Coal		Western Coal	
	O_2(%)	HC (cor ppm)	O_2(%)	HC (cor ppm)
40	15.	114	13.8	125
60	12.7	54	11.3	18
90	9.7	48	--	--
90	8.7	44	8.8	44

No appreciable differences in UHC emissions were noted between eastern and western coals.

Carbon monoxide emissions are a much more sensitive measure of incomplete combustion than are unburned hydrocarbons. A comparison of CO emissions and carbon carryover can be made. Figure 8 is a plot of % carbon in the outlet flyash of a 160 Klb/steam spreader stoker firing western (Montana) coal. This unit exhibited rather high carbon losses

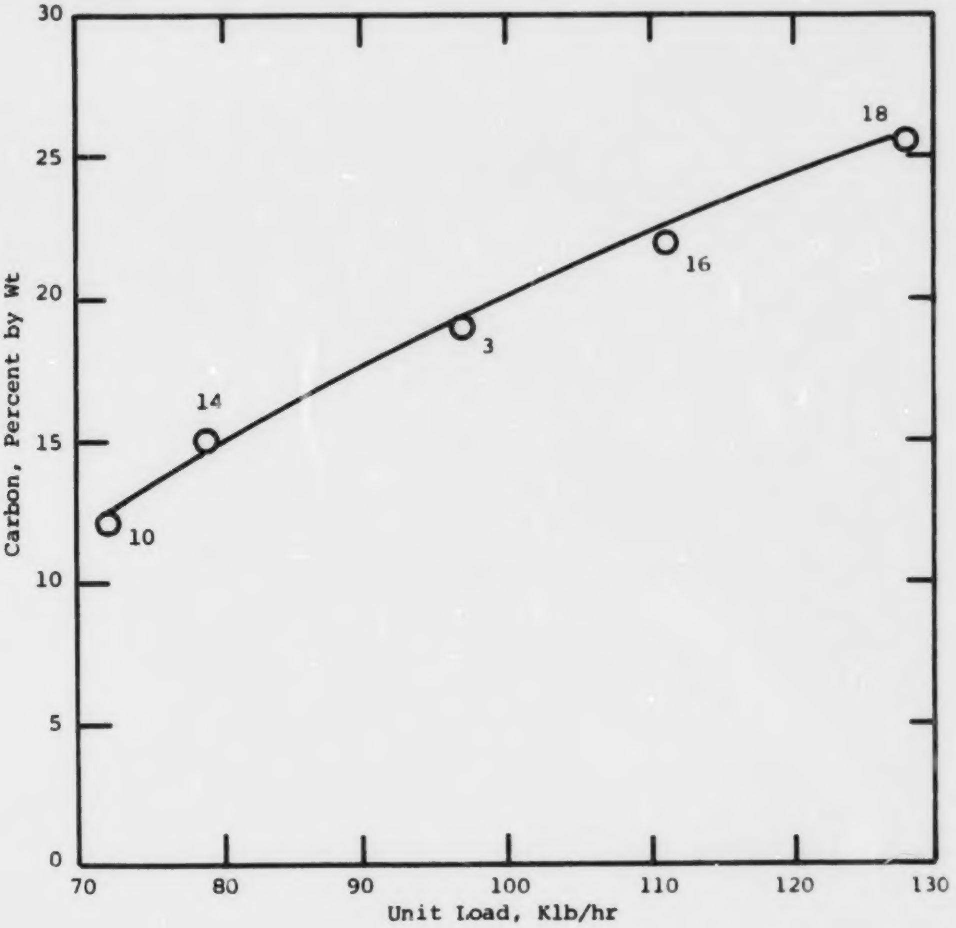

Figure 8. Percent Carbon in Outlet Flyash, 160 Klb/hr Steam Spreader Stoker (Willmar Unit 3), Western Coal.

which increased with unit load. The carbon losses on eastern coal were even larger than for the western coal. However, the point to be made here is that by measuring the carbon monoxide emissions, an indication of the other combustible losses can be gained. The CO emissions for the same tests are shown in Figure 8.

4.2.4 Particulates. For a given ash content in the coal, the quantity of particulate matter in the flue gas from stoker-fired boilers depends primarily upon the amount of burning that takes place in suspension or on the grate. Table VIII gives an average flue gas particulate loading from both eastern and western coals as measured before the control device for the stoker types tested.

TABLE VIII

AVERAGE FLUE GAS PARTICULATE
LOADING FOR EASTERN AND WESTERN COALS

Stoker Type	Klb/hr Capacity	Average Particulate Loading, lb/10^6 Btu Eastern Coal	Average Particulate Loading, lb/10^6 Btu Western Coal	Comments
Spreader Traveling Grate	100	1.643	1.435	13% lower than eastern
Traveling Grate	60	0.783	0.428	45% lower than eastern
Vibrograte	45	0.649	0.488	25% lower than eastern

Spreader stokers with the greater suspension burning have from two to three times the particulate loading of the traveling grate and the vibrograte stokers. The western coal test results showed a lower particulate loading than the eastern coal.

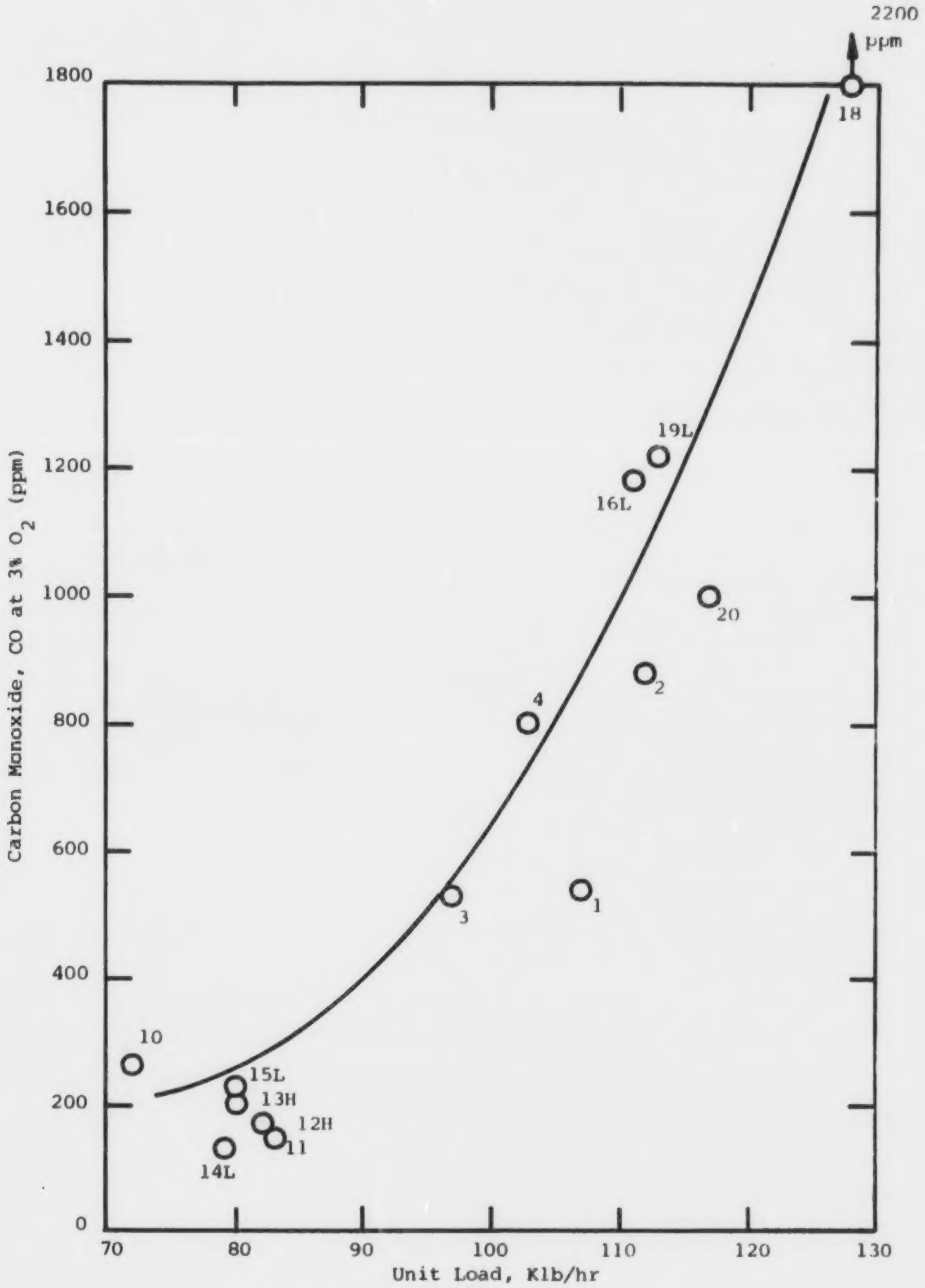

Figure 8. Carbon Monoxide Emissions versus Load (Willmar Unit 3), Western Coal, 160 Klb/hr Steam Spreader Stoker.

5.0 CONCLUSIONS

This study has shown that western subbituminous coals can be substituted for eastern bituminous coals as an industrial boiler fuel. The western coals are compatible with industrial coal-fired units of current design. Two unit types of older design (underfed and traveling grate stokers) were found to have difficulty burning western coal. Some cases have been noted where the maximum load capacity of the boiler had to be limited. This problem can be eliminated by predrying the coal.

Western subbituminous coals were found to be superior to eastern coals in terms of SOx, NOx, particulate, and unburned hydrocarbon emissions. The western coals could be fired at lower excess air and exhibited substantially lower combustible losses than eastern coals.

The size of delivered western coal proved to be a problem in most of the stoker-fired units tested. The coal generally had too large a percentage of fine coal which resulted from the poor weathering characteristics of western coals.

Stoker performance on western coal could be improved if the coal were sized local to the point of use so that delivery distances could be reduced to about 200 miles.

This study has defined the operational parameters that must be followed in order to successfully burn western coal in industrial-sized stokers and pulverized coal units. Excess O_2 and carbon monoxide monitors for combustion control would improve overall industrial boiler performance on both eastern or western coal. These controls are necessary since many times the margin of success can be as small as ± 0.5% excess O_2 in the flue. For the most part, present instrumentation does not provide sufficient precision in combustion control. Operator training and education must go hand-in-hand with improved controls for successful western coal firing.

FACTORS AFFECTING ESP PERFORMANCE ON WESTERN COALS AND EXPERIENCE WITH NORTH DAKOTA LIGNITES

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FACTORS AFFECTING ESP PERFORMANCE ON WESTERN COALS AND EXPERIENCE WITH NORTH DAKOTA LIGNITES

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ABSTRACT

This paper attempts to delineate some of the special characteristics of Western coals that affect the performance of powerplant ESPs. The absence of significant quantities of SO_3 to act as a natural conditioning agent in the flue gas from the low-sulfur Western coals is somewhat offset by their generally higher moisture contents. The alkali constituents, major components of fly ashes from Western coals as contrasted with Eastern or Australian coals, act in various ways to influence ESP performance.

Considerable research into the effect of chemical components on fly ash resistivity has shown sodium to play a significant role in determining resistivity and, hence ESP performance. The ramifications of sodium for ESPs, both as a naturally occurring resistivity reducer and as a possible conditioning agent, are considered.

The Grand Forks Energy Research Center of the Energy Research and Development Administration is designing and constructing a 75-lb/hr pulverized coal-fired furnace and an associated pilot-scale ESP. The objective of the project will be to burn small quantities of a given coal to generate a representative fly ash and flue gas for comparative studies of resistivity and pilot-scale ESP performance. The design and development of the equipment, and its possible usefulness as a design and research tool, are discussed.

A compilation and discussion of ESP design and operating experiences on commercial ESP units at plants burning North Dakota lignites is presented.

FACTORS AFFECTING ESP PERFORMANCE ON WESTERN COALS AND EXPERIENCE WITH NORTH DAKOA LIGNITES

INTRODUCTION

With the increased use of the low-sulfur Western coals for power production has come the realization that previously reliable ESP design parameters do not work well on such coals. The success of ESPs installed on powerplants burning these coals has been mixed, resulting in a wide range of bidder design specifications for a given application. The need for adequate criteria for ESP design on such coals has spurred extensive research toward that end, with some success. This paper will attempt to review the characteristics of Western coals that have been shown to affect ESP performance and discuss their possible usefulness as design indicators. Experience with ESPs on North Dakota lignites will also be described.

PROPERTIES OF WESTERN COALS AFFECTING ESP PERFORMANCE

Coal mined in the Western U.S. generally tends to differ from Eastern coals by having lower sulfur content, higher moisture content, and ash containing a significantly higher proportion of the major alkali and alkaline-earth elements (Na, Mg, and Ca). The coal constituents which are generally accepted as acting as "natural" conditioning agents for ESP operation are moisture and sulfur. The ash chemical analysis is mainly reflected in the resistivity of the collected fly ash. The complicated nature of the ESP/flue gas/fly ash system certainly allows for interactions between effects.

Sulfur

The low sulfur content of most Western coals precludes any significant conditioning effect due to SO_3 in the flue gas. Based on 51 determinations of flue gas SO_3 in four plants burning lignite, only traces of SO_3 could be detected.1/ It appears that for coals with significant quantities of alkali constituents, any SO_3 that is formed reacts quickly with the alkali in the fly ash to produce sulfates. These results indicate a possible area of concern for those considering SO_3 conditioning for fly ashes containing large quantities of calcium. Dismukes reported that a greater increment of SO_3 addition was required to lower measured resistivity to about 10^{10} ohm-cm for a highly basic ash than for a neutral ash.2/ This is most likely due to the formation of calcium sulfate, a high resistivity material.

Moisture

The moisture content of Western coals ranges from around 5% for bituminous coals to over 35% for lignites. A great deal of interest in the effect of coal moisture content on ESP performance has been generated by the excellent operating experience with ESPs on North Dakota lignites. All ESPs constructed on lignite-burning powerplants to date have met or exceeded design efficiencies. A fairly wide range of fly ash chemical analysis and measured resistivity is represented, with little apparent operational difficulty experienced. The extent to which the high moisture content is responsible presents an interesting area for study. Aside from the conditioning of the fly ash to lower resistivity in the low temperature region, flue gas water vapor probably functions like any electronegative gas to enhance ESP performance by allowing a higher operating voltage before sparking becomes a problem. Flue gas moisture could also enhance the effectiveness of a given level of SO_3 addition for ESP conditioning.

The results of a series of three preliminary pilot-scale ESP tests run recently at the Grand Forks Energy Research Center (GFERC) are shown in Figure 1. The tests were run by feeding a high resistivity fly ash into the flue gas at a point upstream of the pilot-scale ESP. The flue gas was generated by the combustion of a Western bituminous coal, with the base moisture level taken as that from the moisture in the coal. Water was injected into the flue gas to provide higher increments of flue gas moisture corresponding to those from subbituminous and lignite coals. The flue gas SO_2 was 270 ppm and the ESP operating temperature 300° F. Due to problems with in-situ resistivity measurements, no data was obtained. The resistivity of this fly ash normally falls in the 10^{12} to 10^{13} ohm-cm range.

As can be seen from Figure 1, the increase in flue gas moisture content from 4 to 8 pct by volume resulted in an increase in effective migration velocity, ω ,from 0.75 to 0.96 fps. These levels correspond to those for bituminous and subbituminous coals, respectively. The increase from 8 to 13 pct flue gas moisture increased ω to 1.06 fps, with the high moisture representing that from a lignite. The effectiveness of an incremental increase in flue gas moisture content would appear to be greater in going from a bituminous to a subbituminous coal than from subbituminous to lignite. The effectiveness of the flue gas moisture change at the lower level is corroborated by results reported by Busby and Darby.3/ They reported the most effective H_2O conditioning with add rates in the range of 0.5-1.0 lb $H_2O/1000\ ft^3$ gas. The levels tested in the GFERC pilot plant were 1.3, 2.7, and 4.3 lb $H_2O/1000\ ft^3$ gas. The coal studied by Busby and Darby was a New South Wales coal which appears to range in moisture contents from 1 to 7.5 pct H_2O in the coal, similar to a Western bituminous coal.4/

Figure 1 also shows the relationship observed between the maximum pilot ESP operating voltage that could be maintained and the flue gas moisture content. A similar set of points was determined to those for ω versus flue gas moisture content. This would seem to indicate that a significant portion of the improvement could be due to the electronegativity of water vapor. Voltage/current curves generated before fly ash was injected verified that the maximum applied voltage possible increased as the flue gas moisture content increased.

Ash Composition

The fly ash chemical composition is the third major factor influencing ESP performance on Western coals. Chemical composition affects are reflected mainly in fly ash resistivities, although reentrainment, hopper ash removal and conditioning effectiveness can also be related to composition. Work by Bickelhaupt5,6/ the authors,7,8/ and others have related measured fly ash resistivity to chemical composition. The sodium content of the fly ash is generally accepted as the best indicator of resistivity, with resistivity decreasing with increasing sodium as illustrated in Figure 2.[8] Figure 3 shows a predictor for fly ash resistivity as a function of fly ash chemical analysis generated by computer regression analysis of data obtained on 35 fly ashes from Western coals.[8]

The conduction mechanism for fly ash is currently described in terms of sodium (or Li or K) ion migration through the outer or surface layers of the glassy fly ash particles. The development of this picture of charge transfer through the fly ash layer by Bickelhaupt[5,6] provides an excellent means of understanding conduction for most fly ashes. However, experience with high-sodium fly ashes and with conditioning by admixture of pure sodium compounds by means of injection at low temperatures indicate that very fine particles of sodium compounds serve to enhance conduction without entering into the glassy phase. Resistivity measurements on such materials indicate that the effect holds up at temperatures up to 800° F as illustrated by the data in Figure 4. This effect of sodium may well be twofold, with the sodium compounds enhancing surface conditioning, and also providing lower resistivity conduction paths through the collected fly ash layer via the very fine sodium materials.

In high-sodium coals, a significant proportion of the sodium upon volatilizing in the flame does not interact with the glassforming elements as a flux. Rather, it reacts with SO_3 to form very fine Na_2SO_4 particles.[9] This formation of very fine particles is partially verified by experience with ESPs on North Dakota lignites, where downstream hopper catches are usually significantly enriched in sodium content, as illustrated in Table 1.

CONDITIONING BY ADDITION OF SODIUM MATERIALS

The only extensive study of sodium conditioning made to date on a full-scale boiler involved injecting sodium into the burner with the coal. White reported on tests utilizing this technique with Na_2CO_3, Na_2SO_4, and NaCl.[10] Each of the three proved to be effective as conditioners, with economics and material availability the major factors in determining feasibility. This method has two drawbacks from an operational standpoint. First there is the potential for increased ash fouling in the superheater region due to the fluxing action of sodium. The problem is well described by the work of Tufte, et.al., [9,11] and could become especially serious where coals with high ash contents are burned. Secondly, a significant portion of the injected sodium may not reach the ESP due to deposition upstream in the heat transfer areas, ductwork and air heater, with the extent of the loss dependent on boiler design. This approach does lend itself to powerplant operation, since mixing the material with the coal during pulverization is reasonably simple.

Tests at GFERC have indicated a conditioning effect for sodium materials injected in dry powder form ahead of the pilot-scale ESP. Previously reported work indicated a significant reduction in fly ash resistivity and improvement in pilot-scale ESP performance when 1 pct or less of previously collected high sodium fly ash (7.6 pct Na_2O) was injected into the flue gas stream ahead of the ESP[8]. Subsequent work showed similar results when comparable quantities of pulverized Nahcolite, a natural-occurring form of $NaHCO_3$, were injected ahead of the ESP. In both cases the flue gas temperature was 400° F at the point of injection. The results are shown in Figures 5 and 6. In both series of tests, the base high resistivity ash and the additive were premixed and then injected into the flue gas ahead of the ESP by means of a solids feeder. All materials were screened to minus 100 mesh. The percentages of additive shown are the percent by weight of the total ash load injected. Total inlet particulate concentrations were in the range of 2.5 to 3 grains/acf for the tests on addition of high sodium ash and 1.8 to 2.0 grains/acf for the Nahcolite work. All ESP tests were run at nominally 300° F with a flue gas flow rate of about 190 acfm. Table 2 lists the addition rates in terms of the total flue gas, illustrating the small quantities of material required as indicated by these tests. These values are probably lower than the addition rates required for a full-scale application, since the pilot tests approach an ideal delivery and mixing situation.

The results of the pilot-scale tests indicate that dry injection of high sodium materials in a powder form may be useful for conditioning ESPs operating on high resistivity fly ash. Application of this technique on a large scale would require the development of an adequate delivery system. The effectiveness of such a conditioning system would depend on the degree of dispersion of the material in the gas stream and on the fineness of the added sodium material.

The economics of sodium conditioning will depend on the cost and availability of the high sodium materials. It would appear that any material with roughly 10 pct or more sodium (reported as Na_2O) would be a candidate for use. High sodium fly ash is a waste material, and there may be other materials containing significant quantities of sodium which are normally waste products. An additional incentive for using Nahcolite is its ability to remove significant quantities of SO_2 from the flue gas, as shown in the same series of tests. Nahcolite levels greater than required for ESP conditioning are necessary for significant SO_2 removal. It is possible that the two operations could be combined under the proper conditions.

TEST FURNACE AND PILOT-SCALE ESP

The Grand Forks Energy Research Center has been actively involved with projects related to the utilization and combustion of Western coals since about 1963. In 1967, a 75-lb/hr pulverized coal-fired furnace was constructed to study ash fouling on boiler tubes. This facility played an important role in determining that sodium content was the chief factor affecting ash fouling rate.[9,11] A flow diagram of this system is shown in Figure 7.

In 1968 a plate-type pilot ESP was retrofited to the furnace to study ESP performance on fly ash from Western coals. Several aspects of ESP operation on such coals have been studied, most notably the affect of chemical composition on fly ash resistivity, and the sodium conditioning work mentioned earlier.[8]

A major problem with the existing furnace has been its inability to provide a fly ash at the pilot ESP inlet representative of full-scale units in terms of particle concentration, size distribution and chemical composition. To overcome this shortcoming and retain the advantage of convenient size, a new 75-lb/hr furnace is presently under construction at GFERC. A cross-sectional view of the unit is shown in Figure 8.

The main combustion section of the new furnace (labeled A) has an inside diameter of two feet and a length of 10 1/2 feet. This section and the following 12 1/2 feet of transitions and cooling zone (B), will be lined with a castable insulating refractory. All but the transitions and furnace bottom will be water cooled by means of an external shell. Six water-jacketed heat exchange sections (single shell and tube) follow, each six feet in length (C). The heat exchangers are designed to allow outlet temperatures from 300 to 700° F. It is believed that maintaining a velocity greater than 50 fps ($\simeq$ 3 fps in the combustion zone) while minimizing bends and surface area should minimize fly ash loss in the combustion train. An additional design improvement for maintaining the fly ash in the gas stream is the substitution of controlled distribution of tertiary combustion air through cross-axial ports in place of mixing by means of swirling the combustion gases. This will hopefully eliminate much of the fly ash loss from deposition on the walls of the combustion zone due to centrifugal force.

A pilot-scale ESP will be attached to the new furnace, as shown in Figure 8. A tube-type ESP has been selected, primarily because it allows for the study of a greater number of variables in ranges approaching those for full-scale units. For example, the maximum flue gas velocity possible at 300° F, with 2 foot high plates and six inch plate-to-plate spacing as in the present unit, is 3.33 fps with the 200 acfm of flue gas available. An 8-inch ID tube, however, allows the range to be varied from 4 to 8 fps by simply bypassing some of the flue gas. Specific collecting areas will still be uncharacteristically low, but in a unit of this size that appears to be unavoidable. It is felt that a small plate-type ESP is probably no more representative of full-scale units than a tube-type in this size range. Also, our primary goal is not to be able to provide dimensional design data, per se, but to provide a ranking of coals on the pilot-scale ESP relative to other coals on which experience has been gained previously by the ESP industry and power companies. The facility will also allow for studies of conditioning methods and basic research on ESP phenomena, such as fly ash conduction mechanism.

EARLY EXPERIENCE ON WESTERN SUBBITUMINOUS COALS

The need to reevaluate the ESP design criteria normally used for fly ash from Eastern coals when considering Western coals became apparent early in the 1960's. Many of the first ESPs installed on boilers burning subbituminous coals failed to meet guaranteed efficiencies. These were low-temperature units designed with specific collecting areas of 90 to 215 ft^2/1000 actual ft^3 for removals in the range of 87 to 98 pct. Gas conditioning by injection of about 20 ppm SO_2 into the flue gas increased removal efficiencies, but not to the level of the guarantee for all units. One utility was able to overcome a high resistivity problem by lowering ESP operating temperatures by diluting the flue gas with fresh air.[12]

EXPERIENCE ON NORTH DAKOTA LIGNITES

All ESPs now in operation on powerplants burning North Dakota lignites have met or exceeded design removal efficiencies. None of the operating or planned units appear to have any unusual degree of overdesign, as can be seen from the data shown in Table 3. Eight of the eleven installations are retrofit units, with four pc-fired, four cyclone-fired, and three spreader-stoker units shown. These units range from 21 to 440 MW capacity. Flue gas temperatures range from 288 to 418° F, with SCAs from 236 to 375 ft^2/1000 acfm. As the table indicates, all units tested to date have operated at 99 pct efficiencies or better, except for the Heskett No. 2 unit designed for 97 pct removal.

The powerplants listed typically produce fly ashes with Na_2O contents within the range of 1 to 10 pct. None of these fly ashes represent an extremely difficult removal problem for an ESP. It has been suggested that the main reason for the high level of success for ESPs on North Dakota lignites is the high moisture content (roughly 35 pct) of the coal. The variations in flue gas moisture from a bituminous to subbituminous to lignite coal would certainly be of a magnitude sufficient to affect ESP performance as demonstrated by Busby and Darby.3/ In fact the typical differences in flue gas moisture between grades of coal are greater than the 0.5 to 1.0 lb H_2O/1000 ft^3 flue gas range required for effective moisture conditioning as shown by their work. However, based on both laboratory and in-situ resistivity measurements, sodium still appears to significantly affect the resistivity of lignite fly ash in spite of the high flue gas moisture. Reports from lignite burning plants also show measureable effects on ESP performance as a result of changes in coal sodium levels. In spite of the current degree of success, caution should be exercised in making generalizations relative to ESP design on any of the Western coals.

Aside from minor wire breakage and TR set inadequacies, the only major problem experienced on lignite-burning powerplants so far has been pneumatic hopper ash removal. For pc- and cyclone-fired units this problem has arisen where a high-sodium fly ash is being collected, and particularly in the downstream hoppers where the collected ash is generally enriched in sodium. The high-sodium fly ashes tend to be sticky and to hold up in the hoppers. Also, high sodium fly ashes tend to have a low bulk density and to absorb moisture at the lower temperature of the outlet hopper, especially if heaters and insulation are not adequate to compensate for the cold North Dakota winters. The solution to the problem has been to increase the rappers, heaters, and insulation on the collection hoppers and conveying ducts. The problem has thus been solved wherever it has occurred so far. An additional problem has occured in one spreader-stoker installation, where carbon carry-over has burned in the hoppers to form clinkers, when the hoppers were allowed to get too hot. Extremely high sodium fly ash has a very low bulk density and is very light and fluffy in appearance. Collection of such a material may result in rapping reentrainment problems.

SUMMARY

The three main factors affecting ESP performance on Western coals are the coal moisture content, coal sulfur content, and fly ash chemical composition. The relatively high moisture content of some Western coals may compensate to some degree for the absence of flue gas SO_3, due to the low sulfur and high alkali content. Sodium content serves as the best indicator of fly ash resistivity for fly ash from such coals. Any proposed ESP installation for a powerplant burning a Western coal should be analyzed on an individual basis, rather than based on generalizations relative to the grade of coal to be burned. The goal of the ESP program at GFERC is to develope a meaningful set of criteria and test equipment to serve as a basis for ESP design on such units. In spite of a high degree of general success with ESPs on powerplants burning Western coals, there is still research to be done to minimize the uncertainties involved in sizing and selecting an ESP for a new installation. In our experience, the cooperation between the ESP industry, the power industry, and the research facilities working on ESP problems has been very good. Continued cooperation and pursuance of the various approaches being taken will answer many of the remaining questions.

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TABLE 1. - Hopper location versus fly ash sodium content, ESPs on North Dakota lignites

Powerplant	Date	Coal Ash Na_2O, percent	Hopper Fly Ash Na_2O, percent			
			Hopper Location			
			Inlet Row	2nd Row	3rd Row	Outlet Row
Leland Olds #1	1/28/75	5.4	7.2	8.3	--	8.4
Leland Olds #2	5/9/75	6.1	9.4	10.7	--	15.4
Hoot Lake #3	11/20/74	4.1	6.6	--	--	8.6
Big Stone	4/8/76		5.7	9.0	7.8	10.6

TABLE 2. - Rate of addition of high sodium materials as ESP conditioners based on pilot-scale ESP tests

Composition of injected fly ash/conditioner mixture	lb conditioner/ 10^6 cf, flue gas	lb Na/10^6 cf, flue gas	Dust loading (grains/acf)	In-situ Resistivity (ohm-cm)	ω (ft/sec)
100% Colstrip	0	0	--	3.3×10^{13}	0.50
.75% High Na Ash	3.03	0.17	2.8	1.5×10^{12}	0.77
10% High Na Ash	39.34	2.22	2.7	2.7×10^{11}	0.91
30% High Na Ash	122.01	6.88	2.8	6.0×10^{10}	0.92
1% Nahcolite	2.64	0.61	1.8	2.3×10^{11}	0.88
5% Nahcolite	14.45	3.35	2.0	4.1×10^{10}	0.99
50% Nahcolite	134.70	31.19	1.9	6.2×10^{9}	1.07

TABLE 3. - ESP design and test data for powerplants burning North Dakota lignites

Utility Company............	Basin Electric Power Cooperative		Minnkota Power Cooperative		Otter Tail Power Company				Montana Dakota Utilities		United Power Association
Station....................	Leland Olds #1	Leland Olds #2	Milton R. Young #1	Milton R. Young #2	Hoot Lake #2	Hoot Lake #3	Ortonville	Big Stone	Heskett #1	Heskett #2	UPA - Stanton
Location...................	Stanton, North Dakota		Center, North Dakota		Fergus Falls, Minnesota		Ortonville, Minnesota	Milbank, South Dakota	Mandan, North Dakota		Stanton, North Dakota
Boiler capacity........MW..	215	440	235	438	61	79	21	440	25	66	160
Firing method..............	pc	cyclone	cyclone	cyclone	pc	pc	spreader-stoker	cyclone	spreader-stoker	spreader-stoker	pc
ESP vendor.................	Research Cottrell	Western	Research Cottrell	Wheel-abrator	Research Cottrell	Research Cottrell	Research Cottrell	Wheel-abrator	Research Cottrell	Research Cottrell	Research Cottrell
New or retrofit installation.............	Retrofit	New	Retrofit	New	Retrofit	Retrofit	Retrofit	New	Retrofit	Retrofit	Retrofit
Completion date............	11/74	9/75	6/75	5/77	5/72	4/72	6/72	5/75	6/75	6/75	5/76
Flue gas											
........Temperature, °F..	360	373	385	380	330	310	345	288	418	333	350
.....volume, 1,000 acfm..	1,000	2,100	1,170	2,200	280	390	133	2,330	189	452	854
..........velocity, Fps..	5.01	5.00	5.55	5.00	4.23	5.07	4.25	5.25	3.80	4.28	5.17
Specific collection area, ...ft^2/1,000 acfm..	320	267	288	375	252	236	280	355	352	280	235
Number of TR sets..........	16	40	16	32	4	4	4	24	6	10	12
Collecting surface/TR set, ...ft^2..	19,970	14,040	21,050	25,800	17,650	23,075	9,310	34,400	11,100	12,650	16,720
Collecting surface/Hopper, ...ft^2..	2,500	1,755	2,630	1,120	3,530	2,560	2,070	1,120	1,850	2,090	2,785
Inlet loading......gr/acF..	2.30	1.30	1.00	1.0 to 2.7	1.87	2.09	0.97	1.17	2.5 to 4.1	0.3 to 0.6[1]	NA[2]
Outlet loading.....gr/acF..	0.0125	0.0125	0.01	0.006	0.015	0.015	0.0042	0.014	0.0225	0.021	NA
Design efficiency.....pct..	99.50	99.05	99.00	99.40	98.50	98.50	98.90	98.80	99.45	97.00	98.00
Measured efficiency...pct..	99.45	NA	99.82	NA	99.00	99^+	99^+	99.63	0.01 gr/ft^3[3]	0.01 gr/ft^3[3]	NA
Migration velocity, ...cm/sec..	8.26	NA	11.15	NA	9.28	9.9	8.4	8.01	NA	NA	NA

[1] ESP downstream of mechanical collector.
[2] Data not available.
[3] Only outlet loading has been measured to date.

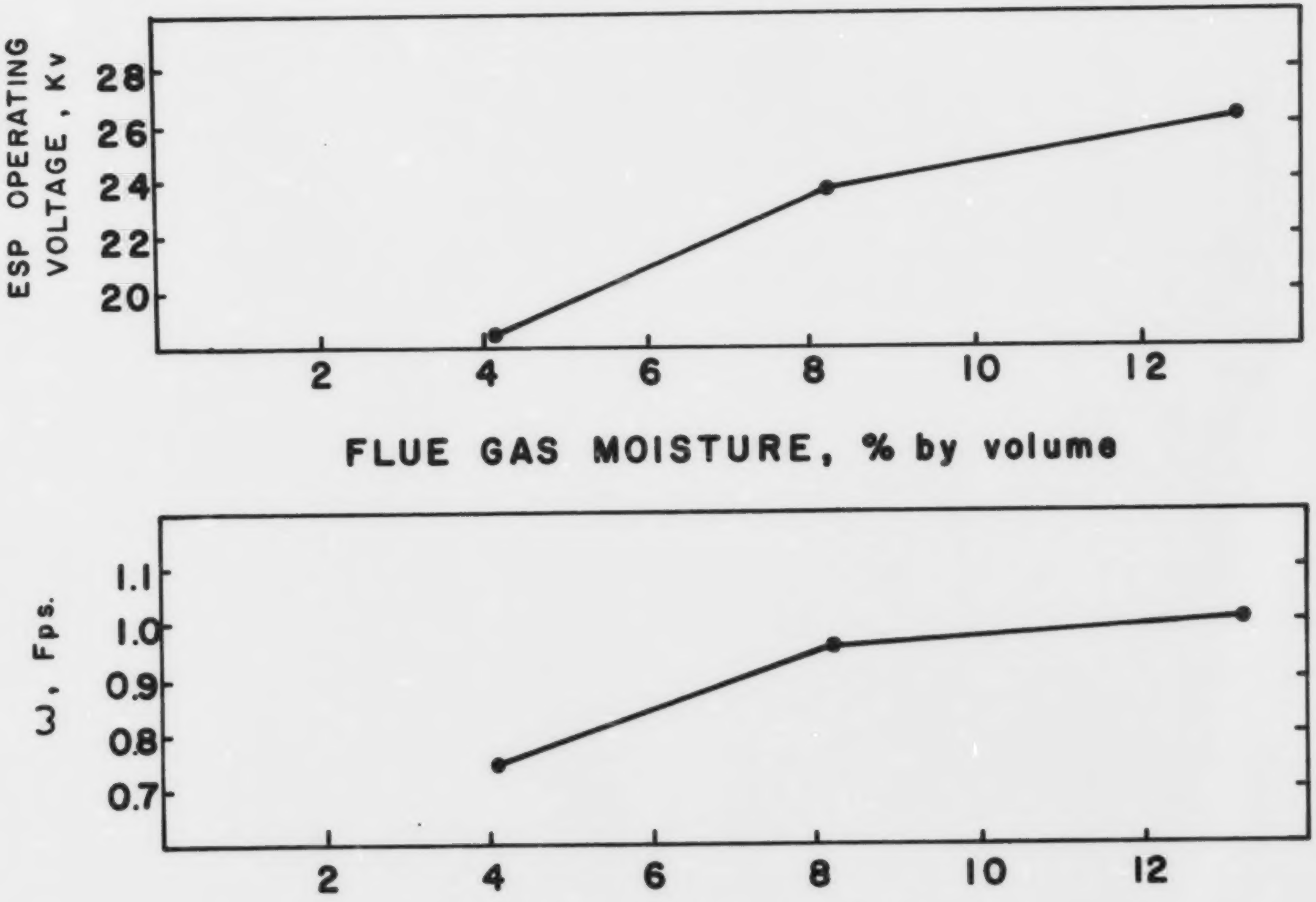

FIGURE 1. - Effect of flue gas moisture on pilot-plant ESP performance.

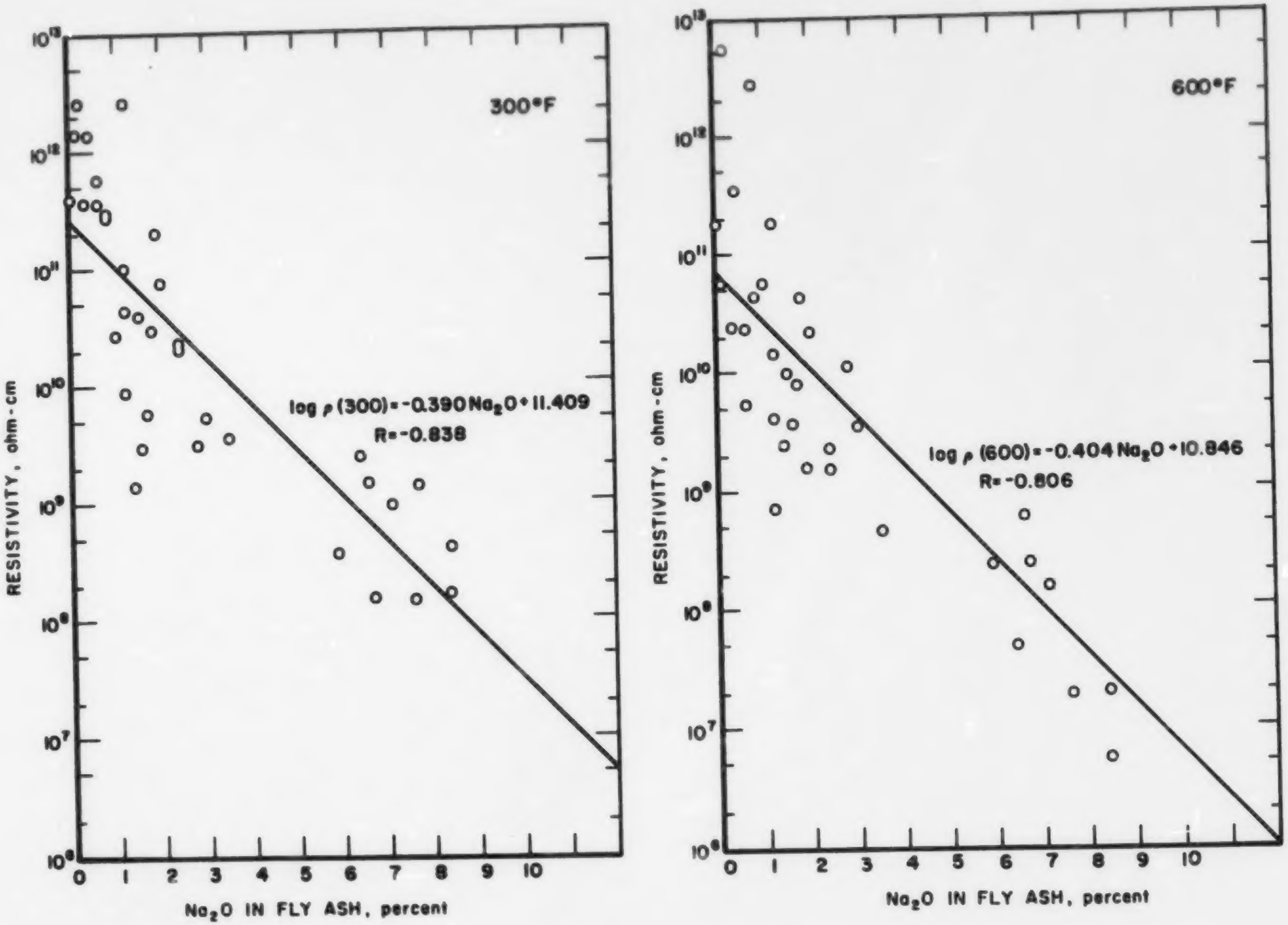

Figure 2 - Laboratory resistivity vs. Na_2O in fly ash at 300 and 600°F.

FIGURE 2. - Effect of fly ash sodium content on measured laboratory resistivity at 300 and 600° F.

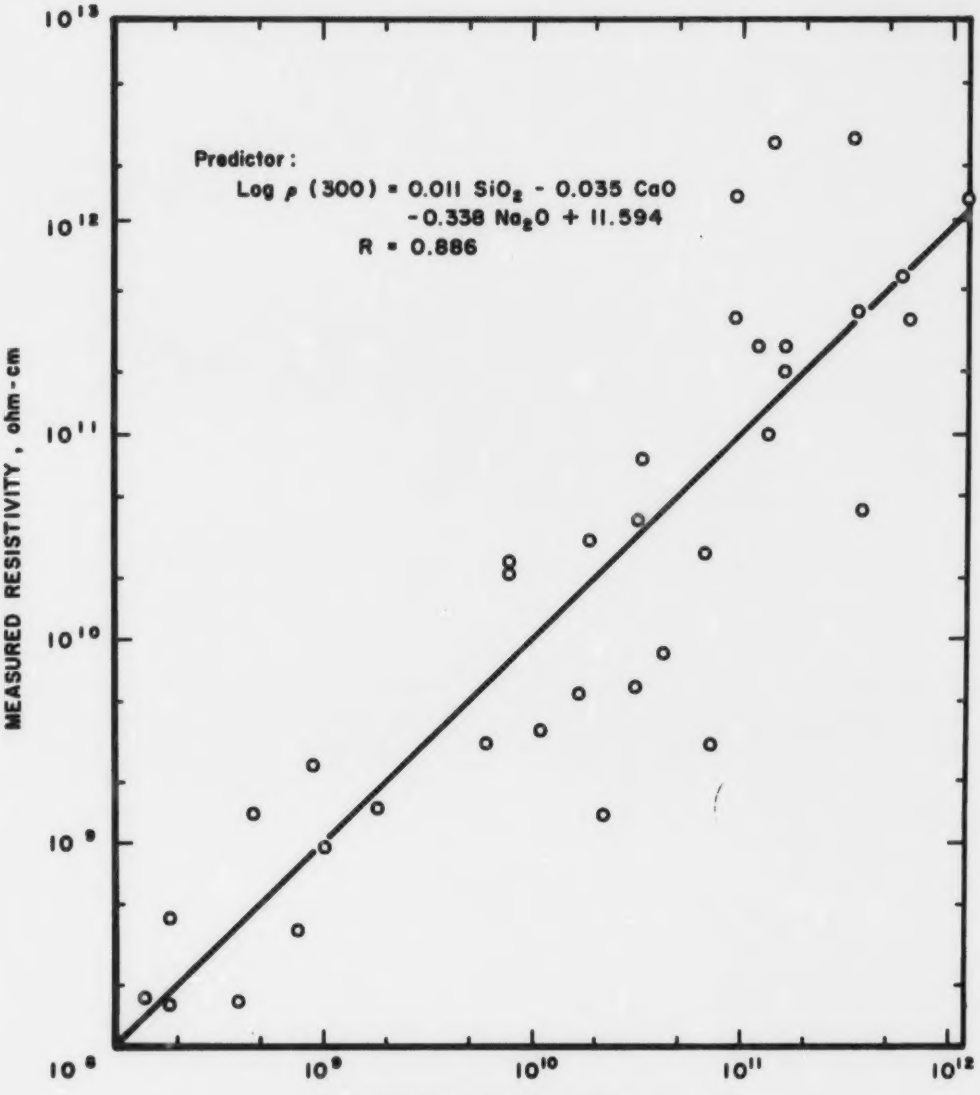

Figure 3 - Measured vs. predicted resistivity at 300°F based on fly ash chemical composition.

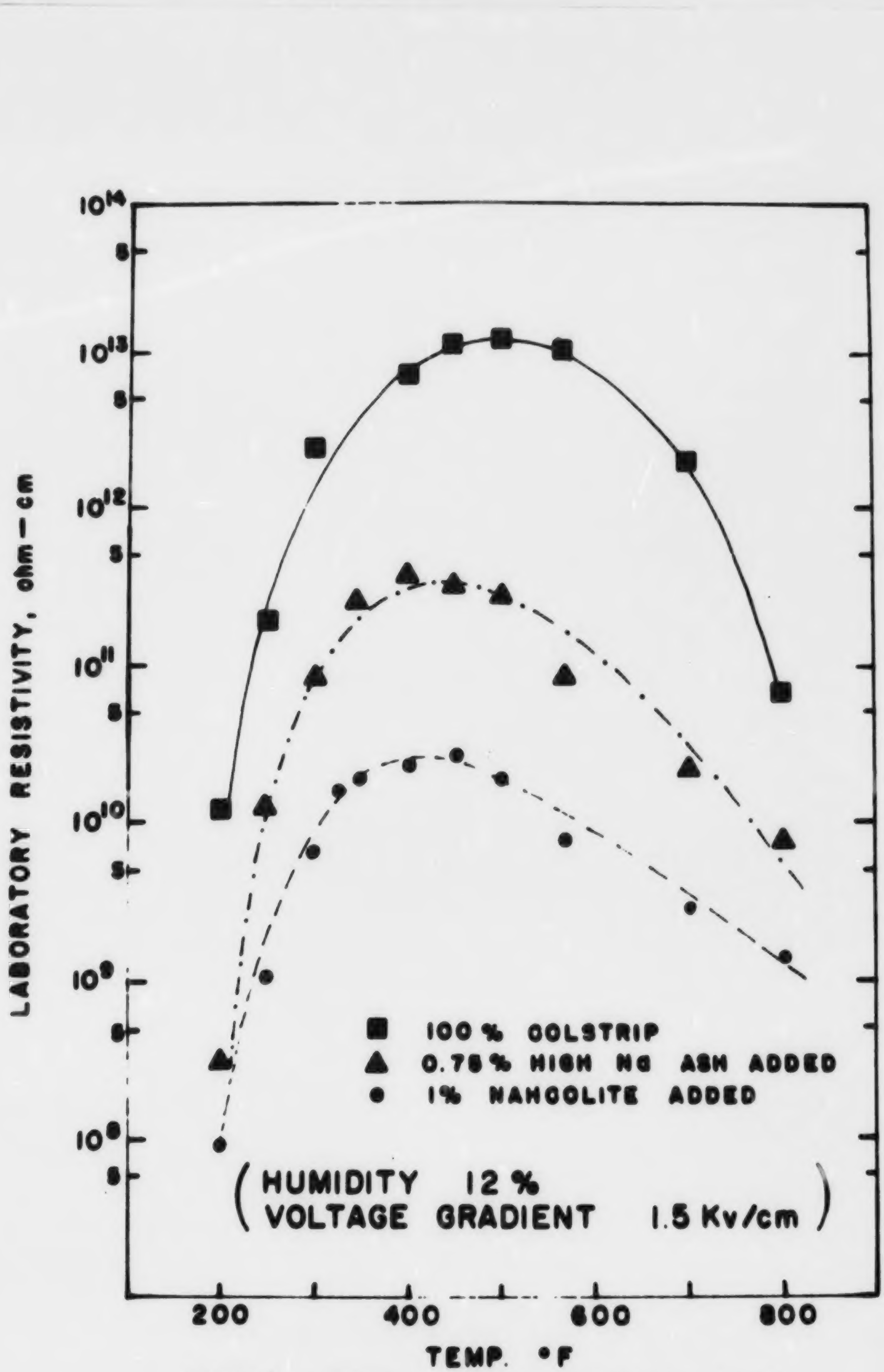

FIGURE 4. - Effect on laboratory resistivity of small quantities of high sodium materials.

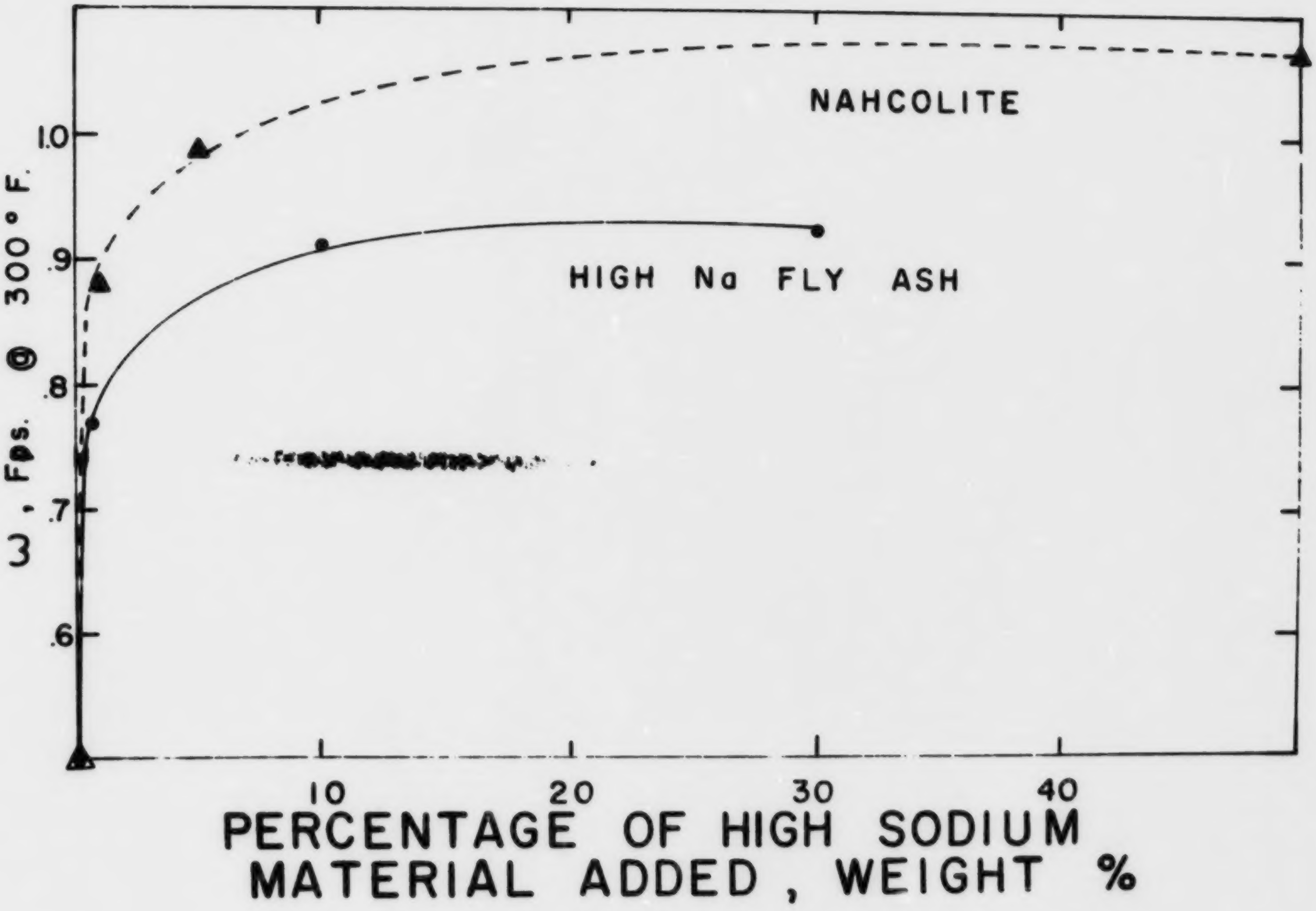

FIGURE 5. - Effect of sodium conditioner on pilot-plant ESP performance.

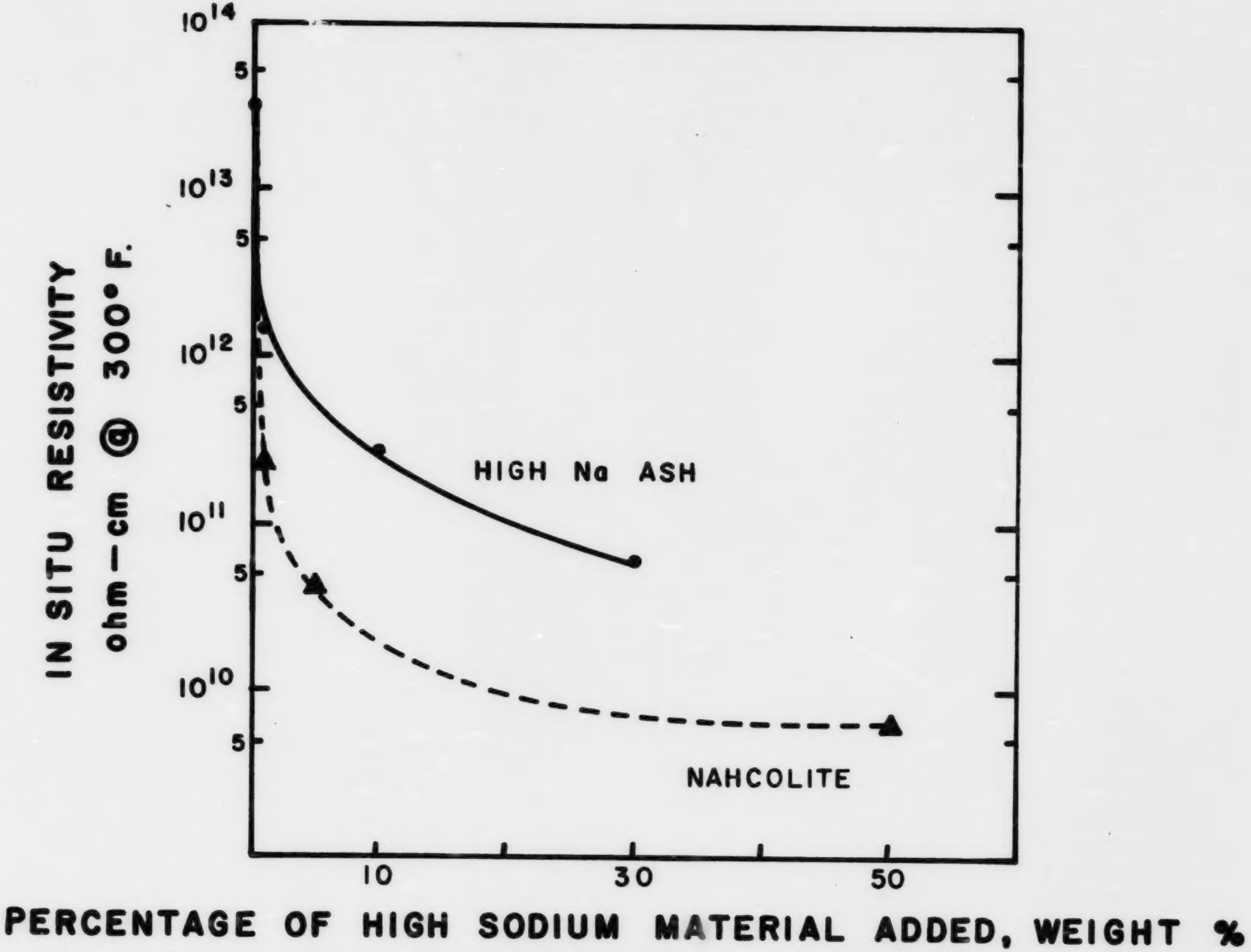

FIGURE 6. - Effect of sodium conditioner on in-situ resistivity.

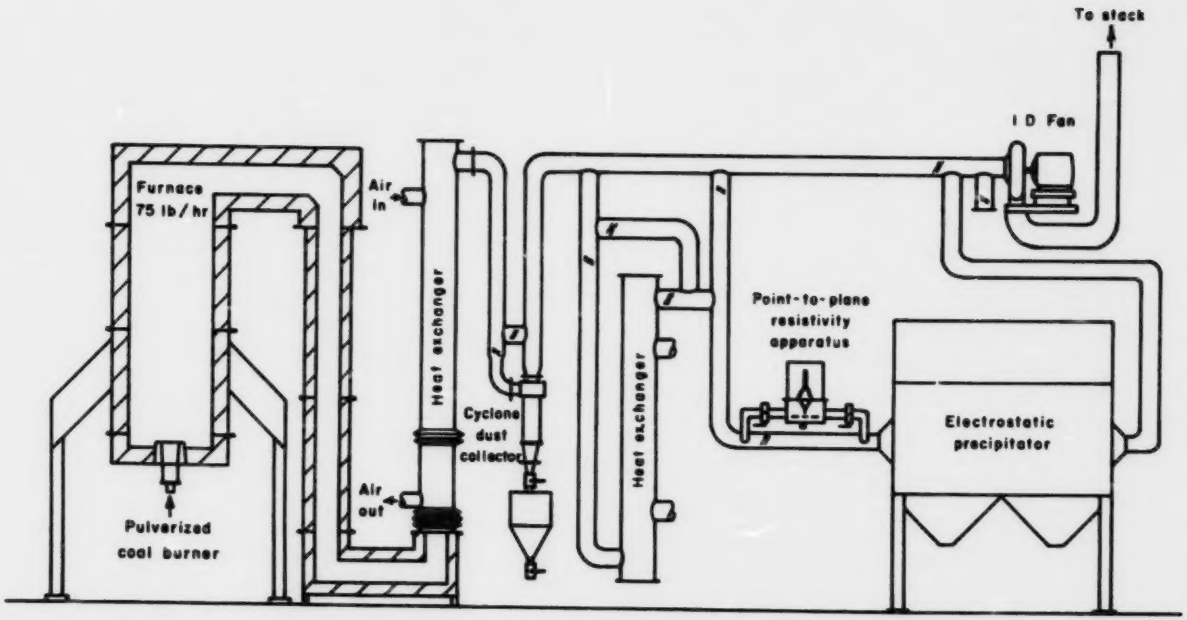

FIGURE 7. - Schematic of pilot plant system used for ESP work to date.

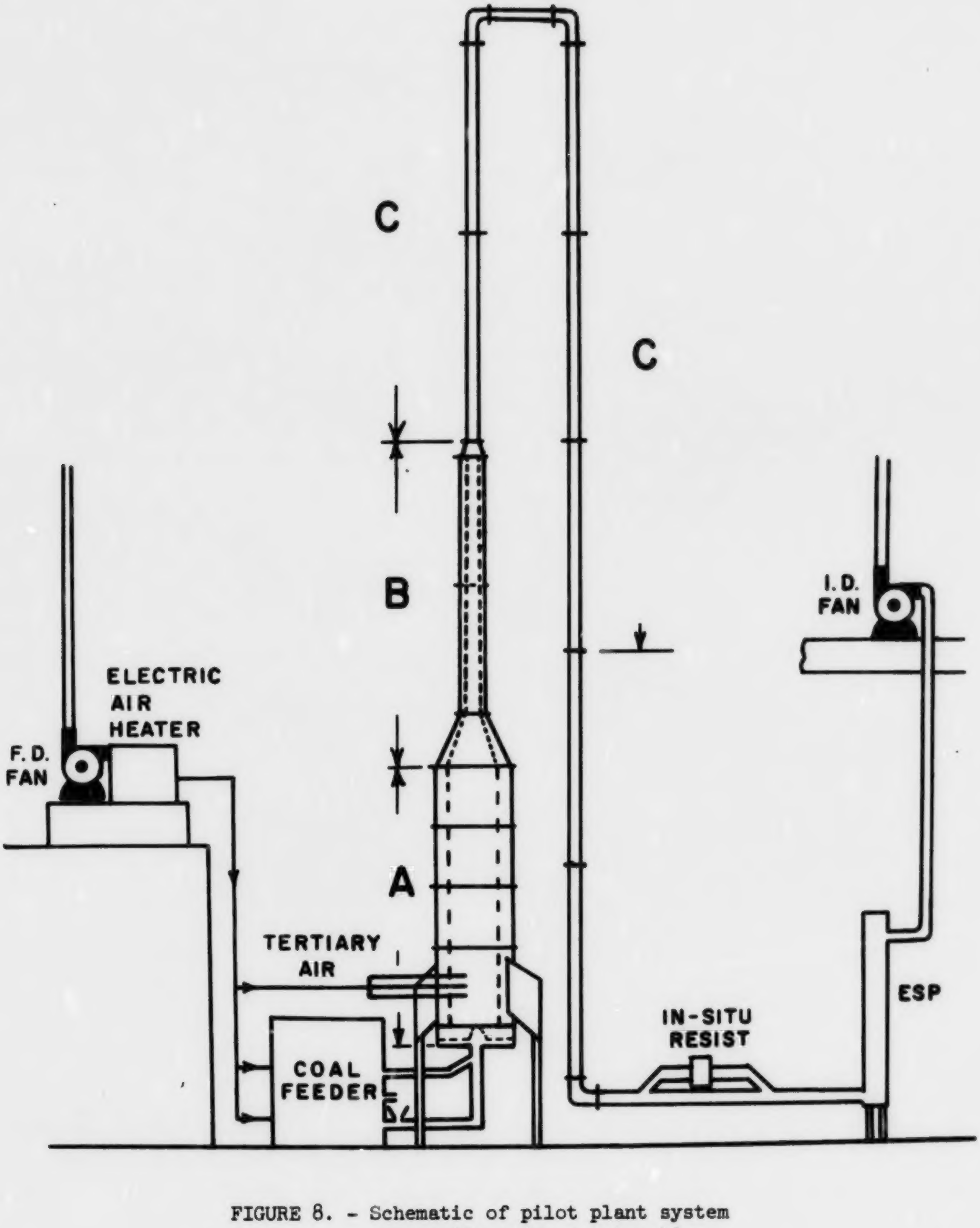

FIGURE 8. - Schematic of pilot plant system currently under construction.

ELECTROSTATIC PRECIPITATOR OPTIONS FOR COLLECTION OF HIGH RESISTIVITY FLY ASH

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ABSTRACT

ELECTROSTATIC PRECIPITATOR OPTIONS FOR COLLECTION OF HIGH RESISTIVITY FLY ASH

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An analysis of the technical options for collecting high resistivity fly ash is presented. The options are a large specific collector area electrostatic precipitator operating at normal temperatures (about 1500°C), a wet electrostatic precipitator, a moderate to large specific collector area electrostatic precipitator operating at either high (about 370 C) or low temperature (about 110°C), a moderate specific collector area normal electrostatic precipitator plus fly ash conditioning. The advantages and disadvantages of each option are discussed. The potential for adverse environmental impact of the conditioning option is also discussed.

ELECTROSTATIC PRECIPITATOR OPTIONS FOR COLLECTION OF HIGH RESISTIVITY FLY ASH

INTRODUCTION

Combustion of low sulfur coal generally produces fly ash with higher resistivity at 150°C than does combustion of high sulfur coal. Thus, fly ash from low sulfur coal is more difficult to collect in electrostatic precipitators than is fly ash from high sulfur coal.

There are several methods of overcoming the influence of high resistivity caused by low sulfur coal. The options are:

1. Provide additional precipitator capacity; i.e., add collector plate area to overcome the poorer performance.
2. Use a wet electrostatic precipitator.
3. Change the flue-gas temperature to achieve a more favorable resistivity.
4. Add fly ash conditioning agents to change the resistivity or otherwise modify electrical conditions in the precipitator.

In deciding which of these options is optimum for a given situation, one must recognize that coals, even from the same mine, vary. This means that the design coal will not be burned all the time. The variation in coal composition, especially of the sodium, iron and potassium content of the coal, Bickelhaupt (1975), will result in variations in fly ash resistivity with resultant variations in electrostatic precipitator performance. Available data indicate that coal composition variations can easily result in day to day resistivity variations of plus or minus a factor of two. The fact that coal composition and thus fly ash resistivity will vary from day to day must be considered in selecting ESP options. The method and costs of dealing with the variation must be considered. Questions such as, "will plate area be added to handle the expected resistivity extremes?" or "will the boiler be operated at reduced load?" must be answered. Regardless of which option is picked, it is essential that the precipitator be well designed, built, operated and maintained. Gas flow distribution must be good to excellent; sneakage and rapping reentrainment losses must be minimized; wire alignment must be good; ash handling must be properly designed; etc.

COLD SIDE ELECTROSTATIC PRECIPITATOR

Cold side electrostatic precipitators (operating at ∿ 150°C) can collect fly ash from low sulfur coal if sufficient collector plate area is provided. The required collector plate area depends on the resistivity of the fly ash and can be as high as 200 $m^2/Am^3/sec$ (∿ 1015 $ft^2/AMCFM$). Specific collector areas in the range of 80 $m^2/Am^3/sec$ to 120 $M^2/Am^3/sec$ (400 $ft^2/AMCFM$ to 600 $ft^2/AMCFM$) are common. High efficiency precipitators with specific collector areas in this range have been built for low sulfur applications.

If the required specific collector area is near the low end of the range cited above, a normal cold side precipitator is likely to be the most economical method of collecting high resistivity fly ash. Some designers feel that in general cold side units are less expensive than hot side units for collection of fly ash from low sulfur, Matts (1975). Other designers believe the opposite, Walker and Fritsch (1976).

The disadvantages of cold side electrostatic precipitators for collecting high resistivity all center around the sensitivity of precipitator performance to small changes in resistivity. This sensitivity to resistivity makes initial design of the precipitator somewhat uncertain, since fly ash resistivity is likely to vary somewhat during the life of the boiler. Once the precipitator is built, normal variations in resistivity can result in performance degradation such that violations of air pollution regulations occur. If the resistivity variation is of a short term nature, emission limits can be met by reducing boiler load. However, if the resistivity change is of long term, such as might be expected if a new coal supply is used to fuel the boiler, long term derating of the boiler or retrofit of additional particulate collection equipment may be required.

WET ELECTROSTATIC PRECIPITATOR

Because the dust layer is continuously washed from the electrodes and the gas is saturated with water vapor, dust resistivity is not a factor in the performance of wet electrostatic precipitators. Thus, wet electrostatic precipitators may be an economically attractive method of collecting high resistivity fly ash.

Whether or not wet precipitators are economically feasible for fly ash collection depends on the cost of purchasing and operating the precipitators. Because wet precipitators must be built of corrosion resistant metaterials, they cost more per square meter of plate area than do dry precipitators. Presently available information indicates that for utility boiler applications wet precipitators are likely to cost 2-1/2 times or more per square meter of collection plate area. If the total collection plate area of the wet precipitator is small enough, the added cost per unit of collector area can be overcome. Performance data, Gooch and McCain (1975), on large plate type wet electrostatic precipitators indicate that specific collection plate area of about 60 m^2/Am^3/sec may be required for high efficiency fly ash collection.

Wet electrostatic precipitators have all the potential water pollution problems that wet scrubbers have. The problems associated with meeting water pollution requirements are likely to increase the cost of wet electrostatic precipitators.

RESISTIVITY MODIFICATION

As discussed in the Cold Side ESP section, precipitator performance becomes extremely sensitive to slight changes in resistivity when the resistivity is high. Under such conditions relatively small change in ash resistivity can lead to excessive particulate emission. Compliance with emission standards can require operation at reduced boiler load with consequent loss of power generation. Therefore, at some point it becomes necessary to consider steps to reduce the resistivity to a more acceptable level.

Resistivity can be changed by operating the precipitator at higher or lower temperature or by adding chemical agents either to the fuel or to the flue gas to condition the fly ash and reduce its resistivity. Although resistivity modification may appear to be more expensive than living with the problem, the advantages of improved confidence in design and lessened sensitivity to small changes in resistivity must also be considered.

HIGH TEMPERATURE OPERATION

Location of the electrostatic precipitator before the air heater allows operation at about 370°C. At this temperature resistivity of most fly ash is in the range 2-4 x 10^{10} ohm-cm or lower. Thus, except for rare problem ashes, hot side precipitators can be designed for relatively high current densities and with confidence that changes in fuel chemical composition are unlikely to affect performance.

Calculations based on the electrostatic precipitator performance model described by Gooch et al (1975) and data reported by Walker and Fritsch (1976) indicate that a specific collector area of at least 67 $m^2/Am^3/sec$ is required for 99.5% particle collection efficiency in a hot side precipitator.

The primary disadvantage of hot side operation is the increase in gas volume due to elevated temperatures. Thus the reduction in SCA for high temperature operation is somewhat offset by the increased gas flow. Table 1 compares calculated collector plate areas for 99.5% particulate removal for various current densities at 150°C with collector plate area required at 370°C on the basis of $m^2/Am^3/sec$ and $m^2/Nm^3/sec$. Note that the hot side precipitator requires a specific collector area of 158 $m^2/Nm^3/sec$.

Problems associated with thermal expansion must be given special attention during the design and construction of hot side precipitators. There have been cases where sufficient care was not taken in dealing with thermal expansion which resulted in poor electrostatic precipitator performance.

Hot side precipitators must be well insulated to reduce heat losses. Insulation and unavoidable heat losses are likely to increase the cost of hot side operation. Walker and Fritsch (1976) present a lucid discussion of design construction and operation of hot side electrostatic precipitators.

LOW TEMPERATURE ELECTROSTATIC PRECIPITATORS

For some cases a 20-50°C reduction in electrostatic precipitator operating temperature may result in resistivities in the acceptable range. The exact nature of the resistivity versus temperature curve at low temperatures is variable and depends on the particular situation. A tenfold or greater variation in low temperature resistivity is common. Thus the option of reducing gas

Table 1. ELECTROSTATIC PRECIPITATOR SPECIFIC COLLECTOR AREA REQUIRED FOR 99.5% PARTICLE COLLECTION FOR VARIOUS CURRENT DENSITIES

Current Density na/cm^2	Temperature °C	Specific Collector Area $m^2/Am^3/sec$	$m^2/Nm^3/sec$
5	150	122	189
20	150	53.1	82.3
40	150	39.4	61
30	370	67	158

Normal conditions 0°C at 1 atmosphere pressure
Calculations based on Gooch et al,(1975)

temperature is subject to verification that the resistivity does indeed decrease with decreasing temperature to a satisfactory level at achievable temperatures. This option is also subject to verification that the rate of corrosion is acceptable; this verification may be possible if the corrosive effect of SO_3 is minimized by the CaO content of the ash. Several precipitator installations have performed satisfactorily at temperatures in the 95-120°C range without undue problems, when sufficient attention was given to the ash-removal system to keep the collected ash hot and to maintain the insulators free of condensed acid.

Achieving a flue gas temperature low enough to overcome the resistivity problem presents problems that must be dealt with on an individual plant basis. If there is sufficient capacity in the air heater, the problem can be handled simply; however, this is not often the case. Whitehead (1971) reported that 120°C is about the minimum temperature that can be achieved even if the air heater has sufficient capacity. An alternative approach to reaching low flue gas temperatures in an existing boiler with insufficient capacity in the air heater or if temperatures below 120°C are required is to increase the flow of secondary air through the air heater and by-pass some of the excess air to the stack. This approach requires additional fan capacity and imposes an energy penalty due to reduction in boiler efficiency, and therefore in most cases is not too attractive.

Berube (1971) presents an excellent discussion of low temperature operation at one plant. For this plant a 0.8% reduction in boiler efficiency was necessary to obtain the required temperature.

In general low temperature operation is sensitive to coal composition. In a situation where coal composition varies over a wide range low temperature may not be practical.

CONDITIONING

The addition of a chemical agent to the fuel or to the flue gas to reduce fly ash resistivity has been successful in some cases. Whitehead (1971) and others report that flue gas conditioning should be considered as an option for most high resistivity applications.

A partial list of presently available or proposed conditioning agents is shown in Table 2.

Table 2. PARTIAL LIST OF CONDITIONING AGENTS

Aluminum Sulfate
Ammonia
Ammonium Bisulfate
Ammonium Sulfate
Hydrogen Chloride
Iron Oxide
Iron Sulfate
Organic Amines
Sodium Carbonate
Sulfamic Acid
Sulfur Trioxide
Sulfuric Acid
Vanadium Oxide

The most commonly used conditioning agent is sulfur trioxide, SO_3. SO_3 is injected into the flue gas either as SO_3 or as sulfuric acid, H_2SO_4. Many of the other agents shown in Table 2 are sold as proprietary agents which may be blends of several of the compounds listed. At the temperatures existing in the flue gas many of the compounds in Table 2 for example, ammonia and sulfur trioxide disassociate into other compounds. Except for SO_3 there is little industrial experience with these conditioning agents some of which have not been demonstrated as being effective.

With the exception of sodium and SO_3 conditioning, the mechanism by which many of the conditioning agents work are unclear. Both sodium conditioning and SO_3 conditioning reduce the resistivity of the fly ash, Bickelhaupt (1974) and Dismukes (1975). Some of the sulfate conditioning agents appear to reduce resistivity probably because SO_3 is one of the products of disassociation, Dismukes (1974). Vanadium and iron oxides are claimed to catalyze the reaction of SO_2 and O_2 to SO_3 and thus increase the quantity of SO_3 present in the flue gas.

The effectiveness of conditioning with some of the agents listed in Table 2 is summarized in Table 3. Based on the available data it appears that properly engineered, operated and maintained conditioning systems using SO_3 as the active agent can improve resistivity limited performance of electrostatic precipitators. The data on most other agents are insufficient to make a general judgment of the effectiveness possible.

There are few data on the fate of conditioning agents after they are injected into the flue gas. It is generally assumed that all the conditioning agent is deposited on the fly ash and then collected in the precipitator, Hall (1973). The available data, although limited, cast doubt on this assumption.

There are more data on the effects of SO_3 than other conditioning agents. Whitehead (1971) reported that in Great Britain conditioning with up to 50 ppm SO_3 resulted in no increase in SO_3 emissions. Dismukes (1975) reported data for two plants that showed from 28 to 80% of the injected conditioning agent was emitted. Published data on SO_3 emissions with and without conditioning for many plants with conditioning systems are lacking. Hall (1973) attributes any increase in SO_3 to poor system design and operation.

Table 3. EFFECT OF CONDITIONING ON EFFICIENCY

	ESP Efficiency or Particulate Emissions	
Agent	Without Injection	With Injection
SO_3	94%	95.2% Dismukes (1975)
SO_3	37.5%	51.4% " "
SO_3	67.3%	77.3% " "
SO_3	69.7%	93.5% " "
Sulfamic Acid	No statistically significant difference, Dismukes (1974)	
Apollo LPA40	$0.349/10^6$J	$0.139/10^6$J, McNinch (1976)
Apollo LPA40	90%	97% Harris (1976)
SO_3	80-88%	95-96% Klipstein (1975)
SO_3	85%	94-95% Cook (1975)

Because SO_3 emitted due to conditioning will form particulate sulfuric acid or particulate sulfates essentially instantaneously upon reaching the atmosphere, it seems reasonable to count the excess SO_3 as particulate sulfuric acid. 1 ppm of sulfuric acid is 5.3 mg/Nm^3 (0.0023 gr/scf^3) of particulate. It is clear that the amount of SO_3 that can escape the stack, if SO_3 is counted as particulate, must be kept at a minimum in order to reduce ambient particulate loading.

Except for data taken by Harris (1976), there are no published data on emissions from other agents. Harris reported that the injection of Apollo Chemical's proprietary conditioning agent LPA40 appeared to result in increased emissions of SO_3, ammonia, surface active agent and other organic compounds. Many of the other proprietary conditioning agents and proposed conditioning agents probably disassociate at flue gas temperatures to produce SO_3 and/or NH_3 -- some of which may be emitted to the atmosphere.

The agents that are claimed to catalyze the oxidation of SO_2 to SO_3 are likely to catalyze SO_2 to SO_3 in the plume as well as in the boiler. Thus, the emission of such an agent to the ambient air may increase the ambient particulate concentration above acceptable limits due to formation of sulfates.

It should be noted that many of the conditioning agents listed in Table 2 are known to be toxic. Thus, an appreciable increase in the emission of these substances may result in adverse health effects -- at least near the power plant.

As stated above there are few data to support or contradict the discussion above. It is clear, however, that toxic substances are being used as conditioning agents and at least some of these agents are emitted to the atmosphere. The impact of these emissions on ambient air quality is unclear.

Conditioning does appear to be an acceptable option for collecting high resistivity fly ash in electrostatic precipitators under some conditions and with some agents. Whenever conditioning is used, careful measurements of all emissions are required to ensure that emissions of toxic substances are not increased. EPA is presently conducting research to determine the overall environmental impact of conditioning agents.

Dismukes (1975) reviewed the capital and operating cost of SO_3 conditioning. His data show that the cost of SO_3 conditioning is sensitive to the method used to produce the SO_3 and the concentration of SO_3 injected.

McNish (1976) reports data on costs of Apollo conditioning agent. Dismukes (1974) presents some data on costs of sulfamic acid conditioning.

CONCLUSIONS

The ideal general solution to the problems associated with collecting high resistivity fly ash from combustion of low sulfur coals does not exist. Utilities who design plants initially for operation using low sulfur coal have several options, one of which is likely to be best for their application. In situations where fuel source and chemical composition is uncertain, hot side electrostatic precipitators should certainly be considered. In cases where coal supply is certain and coal chemical composition is well defined, large cold side electrostatic precipitators may be attractive.

Conditioning is likely to be the least expensive retrofit solution. However, extreme care must be taken in all steps of design, construction and operation of conditioning systems to avoid unacceptable emissions due to the conditioning agents.

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ELECTROSTATIC PRECIPITATOR OPERATING EXPERIENCE WITH LOW SULFUR COAL - JIM BRIDGER POWER PROJECT

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ABSTRACT

BROAD SCALE REVIEWS OF THE ELECTROSTATIC PRECIPITATOR TECHNOLOGY WERE APPLIED TO THE SELECTION AND DESIGN OF THE JIM BRIDGER POWER PROJECT ELECTROSTATIC PRECIPITATORS. INCLUDED WERE INVESTIGATIONS OF MULTI-YEAR EXPERIENCE WITH ELECTROSTATIC PRECIPITATORS BY AUSTRALIAN ELECTRIC UTILITIES, EXTENSIVE PILOT PLANT TESTING, CONSULTATION WITH RECOGNIZED UNITED STATES EXPERTS AND THOROUGH LABORATORY ANALYSES OF COAL AND ASH CHARACTERISTICS INFLUENTIAL IN PRECIPITATOR DESIGNS.

FROM THIS COMPREHENSIVE APPROACH TO ENGINEERING OF AN ELECTROSTATIC PRECIPITATOR INSTALLATION, PACIFIC POWER & LIGHT COMPANY AND IDAHO POWER COMPANY ARE REPORTING ON THE SUCCESS OF ONE OF THE FIRST MAJOR NEW "COLD SIDE" ELECTROSTATIC PRECIPITATORS APPLIED TO A LARGE WESTERN, LOW SULPHUR COAL-FIRED ELECTRIC UTILITY BOILER. THE PRECIPITATORS ARE CAPABLE OF ACHIEVING PARTICULATE REMOVAL EFFICIENCIES IN EXCESS OF 99.3% WITH OUTLET GRAIN LOADINGS LESS THAN 0.02 GRAINS PER ACTUAL CUBIC FOOT PER MINUTE.

ELECTROSTATIC PRECIPITATOR OPERATING EXPERIENCE WITH LOW SULPHUR COAL - JIM BRIDGER POWER PROJECT

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JIM BRIDGER POWER PROJECT

THIS PAPER WILL REVIEW THE ENGINEERING DEVELOPMENT AND EARLY OPERATING EXPERIENCE OF THE ELECTROSTATIC PRECIPITATORS AS APPLIED TO THE JIM BRIDGER POWER PROJECT. INCLUDED WILL BE DISCUSSIONS OF BROAD SCALE INVESTIGATIONS OF THE U. S. AND FOREIGN TECHNOLOGY AND EXPERIENCE, EXTENSIVE PILOT TESTING, SELECTION OF FINAL DESIGN CRITERIA AND RESULTS OF FIRST PERFORMANCE TESTS.

FIRST, AN INTRODUCTION TO THE JIM BRIDGER POWER PROJECT IS APPROPRIATE:

1. LOCATION - WYOMING, 35 MILES ALMOST DUE EAST OF ROCK SPRINGS IN THE SOUTHWESTERN WYOMING COUNTY OF SWEETWATER. THE PLANT IS APPROXIMATELY SIX MILES NORTH OF INTERSTATE HIGHWAY 80 (FORMERLY U. S. 30). THE AREA IS HIGH, 6,800 FEET, THE CONTINENTAL DIVIDE IS NEARBY. THE LAND IS DRY, DESERT SHRUBS AND DESERT GRASSES PREVAIL. THIS IS DESOLATE COUNTRY. ROCK SPRINGS, GREEN RIVER, RELIANCE AND SUPERIOR ARE THOSE NEARBY TOWNS THAT STAND OUT IN THIS WIDE EXPANSE OF THE HISTORIC WEST. THE OLD OVERLAND TRAIL CROSSES THE AREA AS DOES THE MAIN LINE OF THE UNION PACIFIC RAILROAD.

2. Owners - Pacific Power & Light Company and Idaho Power Company, sharing two-thirds and one-third, respectively.

3. Size - Four units, each 500 megawatts. A total of 2,000 megawatts.

4. Fuel - Western coal, low Btu content (9,600 Btu/lb), low sulphur content (0.6%), moderate ash (10%). Coal feed at maximum load is 275 tons per hour each unit.

5. Coal Mine - A strip mine located approximately four miles northeast of the plant site, owned and operated by a subsidiary of the Owners. Coal lands under lease to the Owners from the Union Pacific Railroad and Bureau of Land Management total 18,000 acres. A fuel resource of over 130 million tons with overburden of 120 feet or less exists at this mine. Approximately 125 acres per year will be mined with the Owners dedicated to a full scale program of land rehabilitation.

6. Water Supply - From the Green River by way of a 42-mile pipeline, utilizing controlled release of purchased water from the Fontenelle Reservoir. Water consumption will be 18,700 gpm maximum for four power plant units.

7. Transmission Facilities - Three 345 kv lines from the Jim Bridger switchyard to three major load points in southeastern Idaho.

8. PROJECT STATUS:

A) UNIT 1 WAS PLACED IN COMMERCIAL OPERATION NOVEMBER 30, 1974.

B) UNIT 2 IN COMMERCIAL OPERATION DECEMBER 1, 1975.

C) UNIT 3 IS UNDERGOING START-UP NOW AND IS SCHEDULED FOR COMMERCIAL OPERATION SEPTEMBER 1, 1976.

D) UNIT 4 IS IN THE INITIAL STAGES OF CONSTRUCTION WITH COMMERCIAL OPERATION SCHEDULED FOR DECEMBER, 1979.

THE EARLY SITUATION AND INITIAL TECHNICAL REVIEWS

RELATIVE TO PARTICULATE EMISSION CONTROL, QUALITY OF THE AIR AND THE JIM BRIDGER POWER PROJECT, THESE WERE THE CIRCUMSTANCES PREVAILING AT THE BEGINNING. THE PERIOD IN TIME IS 1970-1971. PUBLIC CONCERNS FOR PROTECTION OF THE ENVIRONMENT, PARTICULARLY AIR QUALITY, WERE MOUNTING. ACROSS THE COUNTRY, STRINGENT STATE AND FEDERAL AIR QUALITY CONTROL REGULATIONS WERE IN VARIOUS STAGES OF "DRAFT," PROMULGATION AND ENFORCEMENT. WYOMING HAD PARTICULATE EMISSION STANDARDS OF 0.18 LB. PER MILLION BTU INPUT (0.05 GRAINS PER ACF) AND A RINGLEMANN #1 OPACITY LIMITATION. PACIFIC POWER AND IDAHO POWER SAW A NEED FOR A NEW POWER PLANT WITH "CLEAR" CHIMNEYS -- CHIMNEY PLUMES THAT WOULD NOT VIOLATE THE BLUE SKY, PRISTINE CONDITION OF THE WIDE OPEN ATMOSPHERIC SPACES OF WYOMING.

A TECHNICAL APPRAISAL AT THIS STAGE EQUATED A "CLEAR" CHIMNEY TO PARTICULATE EMISSIONS NOT EXCEEDING 0.02 GRAINS PER ACTUAL CUBIC FOOT OF FLUE GAS.

WHERE DID THE PARTICULATE COLLECTION TECHNOLOGY STAND AT THIS POINT IN TIME TO SUPPORT A CLEAR CHIMNEY DESIGN? DISMAL AT BEST WAS THE TECHNICAL SITUATION. THERE WERE NO HIGH EFFICIENCY PRECIPITATORS OPERATING SUCCESSFULLY IN THE U. S. IN THE PROCESS OF COLLECTING LOW SULFUR COAL ASH. THOSE PRECIPITATORS IN OPERATION WERE EITHER NOT OPERATING SUCCESSFULLY OR NOT DESIGNED FOR COLLECTION EFFICIENCIES IN THE 99%+ RANGE -- OR CLEAR CHIMNEY RANGE.

TECHNICAL UNDERSTANDINGS OF ASH RESISTIVITY, ASH CONSTITUENTS, MOISTURE, GAS CONDITIONING, TRANSVERSE RAPPING ACCELERATIONS, RIGID VS SUSPENDED WIRE ELECTRODES, "HOT" VS "COLD" WERE STILL DEVELOPING, AT LEAST AS APPLIED TO HIGH EFFICIENCY COLLECTION OF LOW SULFUR, WESTERN COAL ASH PARTICLES. THIS IS ALL TRUE WHEN VIEWED FROM THE BACKGROUND OF UNITED STATES EXPERIENCE. IN EUROPE, INCLUDING GREAT BRITAIN, AND IN AUSTRAILIA, MULTI-YEAR EXPERIENCE HAD BEEN RECORDED IN THE SUCCESSFUL HIGH EFFICIENCY COLLECTION OF ASH PARTICLES FROM COALS SIMILAR TO THOSE EXISTING IN THE WESTERN UNITED STATES.

PACIFIC POWER, IDAHO POWER AND THEIR ENGINEERS, THEREFORE, UNDERTOOK EXTENSIVE INVESTIGATIONS OF PRECIPITATOR EXPERIENCE IN AUSTRALIA (NEW SOUTH WALES AND QUEENSLAND) AND GREAT BRITAIN. A SUMMARY OF IMPORTANT OBSERVATIONS AND CONCLUSIONS RESULTING FROM THESE INVESTIGATIONS FOLLOWS:

1. Australia - With the very thorough and expert assistance of several engineers of the Electricity Commission of New South Wales, plant site investigations were undertaken at several of the Commission's power plants -- Munmorah, Vales Point, Liddell and Tallawara in New South Wales, as well as the Swanbank Power Station in Queensland (Southern Electricity Authority). Here were several large scale, cold side precipitators operating with collection efficiencies in excess of 99% and collecting ash particles from low sulfur coal (1/2% or less). Chimney plumes were not clear. Inlet grain loadings were high, two to three times greater than expected loadings at Jim Bridger. Therefore, 99%+ collection still resulted in outlet grain loadings of approximately 0.1 grain per cubic foot, well within the visual plume range. Also, well within the Australian emission standard of 0.2 grains/cu. ft. existing at that time. The important point here is that low sulfur coal ash was being collected at high efficiency levels on a sustained basis.

The Electricity Commission of New South Wales, in 1971, had a well defined approach to the sizing and specification of new precipitator equipment. The principal element in this approach was thorough pilot testing of the coal to be burned in the new unit. Need for such testing had been demonstrated in the ECNSW experience by an inability to properly size precipitator equipment based solely on examination and comparison of coal and ash laboratory analyses.

FURTHER EXPERIENCE OF THE ECNSW ESTABLISHED THE NEED FOR VERY HEAVY RAPPING FORCES. NOT ONLY IS THE AUSTRALIAN ASH DIFFICULT TO PRECIPITATE, IT IS ALSO DIFFICULT TO REMOVE FROM THE COLLECTING PLATES. AUSTRALIAN SPECIFICATIONS THEREFORE CONTAIN THE REQUIREMENT FOR A MINIMUM TRANSVERSE ACCELERATION OF 50 G'S AT ANY POINT ON THE PLATE. THE GENERAL THEORY IS HEAVY, INFREQUENT RAPPING AS OPPOSED TO LIGHT, FREQUENT RAPPING. AND SUCH A HEAVY RAPPING SPECIFICATION DICTATES A STRONG, RUGGED STRUCTURAL DESIGN OF PLATES AND SUPPORTING COMPONENTS.

ONE OTHER OBSERVATION FROM THE AUSTRALIAN INVESTIGATIONS. TO ASSIST THE ECNSW IN RESOLUTION OF THE SEVERAL BAFFLING ISSUES EXISTING WITHIN THE ELECTROSTATIC PRECIPITATOR TECHNOLOGY, SERVICES OF THREE LABORATORIES WERE ACTIVE IN DEVELOPMENTAL AND RESEARCH WORK. COMMONWEALTH SCIENTIFIC INDUSTRIAL RESEARCH ORGANIZATION, CISRO, HAS BEEN INVOLVED IN COAL ASH PRECIPITABILITY AND PARTICLE SIZE RESEARCH. AUSTRALIAN COAL INDUSTRY RESEARCH LABORATORY, ACIRL, HAS CONDUCTED EXTENSIVE RESEARCH IN THOSE AREAS OF PARTICLE SHAPE AND DISTRIBUTION EMPLOYING PHOTOMICROSCOPY. WOLLONGONG UNIVERSITY COLLEGE HAS CONCENTRATED EFFORTS IN THOSE AREAS OF A BETTER UNDERSTANDING OF ASH RESISTIVITY AND THE ADHESIVE FORCES OF ASH DEPOSITED ON COLLECTING SURFACES.

FROM AUSTRALIA THEN, THESE PERTINENT POINTS:

A) LOW SULFUR COAL ASH WAS BEING COLLECTED AT HIGH LEVELS OF EFFICIENCY, COLD SIDE;

B) PILOT TESTING HAD BEEN DETERMINED TO BE A REQUIREMENT TO CONFIRM A SATISFACTORY DESIGN FOR AN UNKNOWN COAL;

c) HEAVY RAPPING FORCES AND A RUGGED STRUCTURAL DESIGN TO WITHSTAND THESE FORCES WERE CONSIDERED NECESSARY TO KEEP COLLECTING PLATES CLEAN AND PRECIPITATOR PERFORMANCE AT A CONTINUING SATISFACTORY LEVEL.

2. GREAT BRITAIN - ONE OF THE OWNER'S ENGINEERS VISITED THE STAYTHORPE "B" POWER STATION OF THE CENTRAL ELECTRICITY GENERATING BOARD IN ENGLAND. AGAIN, THE PRINCIPAL INPUTS TO THE JIM BRIDGER PRECIPITATOR DECISION FROM THIS VISIT WERE:

A) STAYTHORPE PRECIPITATORS WERE OPERATING SATISFACTORILY AT COLLECTION EFFICIENCIES IN EXCESS OF 99%, TREATING ASH FROM LOW SULFUR COALS.

B) THE SPECIFIC PRECIPITATOR DESIGN INCLUDED HEAVY RAPPING.

MANUFACTURER PREQUALIFICATION

IN MID-1971, THE JIM BRIDGER OWNERS ISSUED TO CANDIDATE ELECTROSTATIC PRECIPITATOR MANUFACTURERS A LETTER REQUESTING INFORMATION TO BE USED IN A VENDOR PRE-QUALIFICATION SURVEY. THIS PRE-QUALIFICATION LETTER REQUESTED:

1. AN INSTALLATION LIST OF ELECTROSTATIC PRECIPITATORS SUCCESSFULLY OPERATING ON UNITS WTH LOW SULFUR COAL (1% OR LESS) AND EFFICIENCIES OF 99% OR GREATER.

2. A STATEMENT OF APPROPRIATE PRECIPITATOR TESTS AND ANY EVIDENCE OF SUSTAINED SATISFACTORY PERFORMANCE.

3. THE APPROACH USED IN DETERMINING THE SIZE FOR THE ABOVE UNITS AND A STATEMENT RELATIVE TO MANUFACTURERS EXPERIENCE AND USE OF PILOT PRECIPITATORS.

4. A STATEMENT OF MANUFACTURERS PROPOSED METHOD OF SIZING THE JIM BRIDGER PRECIPITATORS.

5. INFORMATION AS TO THE AVAILABILITY OF A THREE-STAGE PILOT PRECIPITATOR.

THE VENDOR QUALIFICATION CRITERIA HERE WAS TO LIMIT PROPOSALS TO THOSE PRECIPITATOR MANUFACTURERS WITH DIRECT EXPERIENCE AND SUCCESS IN THE DESIGN AND FURNISHING OF EQUIPMENT TO OPERATE ON LESS THAN 1% SULFUR COAL ASH AND AT A COLLECTION EFFICIENCY OF 99% OR GREATER. ADDITIONALLY, THE CRITERIA SEEKS TO ESTABLISH AN UNDERSTANDING OF PRECIPITATOR SIZING TECHNIQUES AND THE MANUFACTURERS POSITION WITH REGARD TO THE APPLICATION AND USE OF PILOT TESTING. PILOT TESTING WAS GOING TO BE A MAJOR PART OF THE JIM BRIDGER PRECIPITATOR DEVELOPMENT PROGRAM.

ELECTROSTATIC PRECIPITATOR SPECIFICATION

FROM THESE EARLY INVESTIGATIONS COME SEVERAL UNIQUE FEATURES OF THE BIDDING SPECIFICATION FOR THE JIM BRIDGER ELECTROSTATIC PRECIPITATORS:

1. PROPOSALS WERE TO BE BASED UPON A COMPLETE DESIGN FOR PRECIPITATORS WITH A BASE SPECIFIC COLLECTING AREA OF 400 SQ. FT./1,000 ACFM.

2. Unit prices were to be quoted for subsequent adjustments in specific collecting area, number of electrical sets and changes in duct configuration.

3. Separate gas paths were specified with the specific requirement that any one gas path could be removed from service. With one gas path out of service, the precipitator was specified to treat full gas flow and not exceed guaranteed maximum particulate emission rates. Complete isolation capability of each gas path was called for, electrically and mechanically, so that maintenance within and upon the out of service gas path could be accomplished.

4. Guaranteed performance at an emission rate of 0.02 grains/acf (equivalent to 99.33% collection efficiency based on maximum calculated inlet grain loading). The maximum particulate emission rate was to be guaranteed when firing coals within the range of 33 separate core hole sample analyses representing the entire coal field.

5. A minimum collecting plate rapping acceleration of 50 g measured normal to the surface was specified.

6. Provision for a spare empty section (20% additional surface) was specified.

7. Flue gas conditioning would not be permitted.

8. A COMPLETE DESCRIPTION OF MODIFICATIONS, ADJUSTMENTS AND/OR ADDITIONS CONTEMPLATED TO CORRECT PERFORMANCE DEFICIENCIES IN THE EVENT GUARANTEES WERE NOT MET WAS A SPECIFIED REQUIREMENT.

9. GAS FLOW DISTRIBUTION MODEL TESTING WAS ALSO A SPECIFIED REQUIREMENT.

10. TO ESTABLISH FINAL DESIGN PARAMETERS, SPECIFIC COLLECTING PLATE AREA, ELECTRICAL CHARACTERISTICS, ETC.; PILOT PRECIPITATOR TESTS WERE SPECIFIED. THE SUCCESSFUL VENDOR WAS TO SUPPLY:

    A) PILOT ELECTROSTATIC PRECIPITATOR, COMPLETE, WITH AT LEAST THREE ELECTRICAL SECTIONS TO BE USED AT A POWER STATION (AT THIS TIME NOT YET IDENTIFIED) FIRING JIM BRIDGER COAL.

    B) TECHNIQUES FOR CORRELATION OF PILOT TEST RESULTS WITH JIM BRIDGER DESIGN AND PERFORMANCE ELEMENTS.

    C) ALL TEST INSTRUMENTATION.

    D) TEST SUPERVISION AND PERSONNEL.

    E) DESIGN OF CONNECTING INLET AND OUTLET DUCTWORK.

    F) CHEMICAL ANALYSES, SIZE DISTRIBUTION TESTS AND RESISTIVITY TESTS FOR BOTH PARTICULATES AND COAL AS APPROPRIATE.

G) Test procedures, including measurement techniques were to be developed jointly, owners and manufacturer.

11. Performance tests of the Jim Bridger units were specified as follows:

A) Preoperational tests - a physical inspection and cold air flow distribution test to verify proper construction.

B) Period No. 1 - starting upon initial operation and ending with written notification that equipment is ready for satisfactory operation. A period of 1,500 hours or less of on load operation, allowing for manufacturers operational tests and adjustments as required.

C) Period No. 2 - starting upon receipt of notification that equipment is ready for satisfactory operation and extending for 2,000 hours of on-load operation. No alteration of control settings, no adjustments, no modifications, no additions would be allowed.

D) Performance Test "A."

E) Period No. 3 - starting at the successful completion of Test "A" and extending for a period of 4,000 hours of on-load operation. Again, no control setting alterations, no adjustments, no modifications, no additions would be allowed.

F) PERFORMANCE TEST "B" - SUCCESSFUL COMPLETION OF TEST "B" WILL CONSTITUTE THE DATE OF PLACING THE EQUIPMENT INTO COMMERCIAL OPERATION.

DURING THIS PERIOD OF TIME IN THE JIM BRIDGER PRECIPITATOR DEVELOPMENT PROGRAM, THERE WERE CONTINUING INVESTIGATIONS AND TECHNICAL INPUTS. THE AUSTRALIAN LABORATORIES CISRO, ACRIL AND WOLLONGONG UNIVERSITY COLLEGE HAD COMPLETED ANALYSES AND STUDY OF JIM BRIDGER COAL SAMPLES. MANUFACTURERS ACTIVE IN THE PROGRAM WERE CONTRIBUTING FURTHER TECHNICAL UNDERSTANDINGS. IN ADDITION TO BECHTEL CORPORATION, ENGINEERS FOR THE PROJECT, CONSULTATIONS WITH SEVERAL TECHNICAL EXPERTS WERE CONTINUING. DR. HARRY J. WHITE, RETIRED PORTLAND STATE UNIVERSITY PROFFESSOR; INDIVIDUALS FROM SOUTHERN RESEARCH INSTITUTE AND MR. KEN WATSON, ELECTRICITY COMMISSION OF NEW SOUTH WALES, ALL WERE INVOLVED DEEPLY IN THE PROGRAM DEVELOPMENT AT THIS STAGE. TWO INTERESTING CHANGES WERE MADE IN THE ELECTROSTATIC PRECIPITATOR SPECIFICATION AS A RESULT OF THESE FURTHER INVESTIGATIONS.

THE COLLECTION EFFICIENCY GUARANTEE OF 99.33% AND THE EMISSION RATE GUARANTEE OF 0.02 GRAINS/ACF WERE TIED TO THOSE COAL FIELD CORE SAMPLES WHOSE ANALYSES INDICATED AN ALKALI-SULFATE INDEX (ASI) EQUAL TO OR GREATER THAN THE ALKALI-SULFATE INDEX OF THE PILOT TEST COAL (ASI_p). THE ALKALI-SULFATE INDEX IS AN EMPIRICAL TECHNIQUE IN THE ASSESSMENT OF ASH RESISTIVITY AND THEREFORE ASH PRECIPITABILITY. IT IS NOT THE PURPOSE OF THIS PAPER TO DISCUSS THE TECHNICAL VALIDITY OR DETAILS OF THE ALKALI-SULFATE INDEX OUTSIDE THE INFLUENCE ON THE FINAL JIM BRIDGER ELECTROSTATIC PRECIPITATOR SPECIFICATION. IN THE SPECIFIC CASE OF JIM BRIDGER, THE RESULT WAS THAT THE EMISSION GUARANTEES (0.02 GRAINS/ACF AND 99.33%) WOULD APPLY ONLY DURING THE FIRING OF COALS WITH AN ASI GREATER THAN ASI_p, INCLUDING COALS REPRESENTED BY 25 OF THE 33 CORE SAMPLES.

Further economic evaluations dictated a prudent modification to the original specification that guaranteed performance be maintained with full gas flow and one gas path out of service. The final specification language reads, "Seller guarantees a collection efficiency of 98.73% with one gas path removed from service with flue gas flow limited to 95% of design,....."

Based on these prequalification criteria and specifications, a joint venture of A. B. Svenska Flaktfabriken and Carborundum Incorporated was selected as the supplier of the Jim Bridger Unit 1, 2 and 3 Electrostatic Precipitators. The date, October, 1971. The total cost - $32,000,000 for three units, equal to $21+ per kilowatt.

PILOT PRECIPITATOR UNIT

Concurrent with the selection of the Jim Bridger Electrostatic Precipitator manufacturer, selection of a power plant site for the pilot testing work, opening of the Jim Bridger coal mine as a fuel source and arrangements for shipment of the coal to the test plant were all underway. Pertinent details of these three phases of the program follow:

1. Test Site - Pacific Power & Light Company's Dave Johnston Plant, approximately 30 miles due east of Casper, Wyoming was selected as the test site. Unit 1 has these statistics: 100 mw; B&W, 1550 psig, 1000°F, single reheat, front fired boiler; B&W EL76 ball mills.

2. Test Coal - From a review of the Jim Bridger coal mine core hole analyses, an area representative of the entire coal field was selected. There

WERE, OF COURSE, PRACTICAL MINING LIMITATIONS IMPOSED ON THIS SELECTION PROCESS. THIRTY THOUSAND TONS WERE MINED FROM THIS AREA, THE FIRST COAL FROM THE JIM BRIDGER FIELD.

3. COAL SHIPMENT AND HANDLING - THE TEST COAL WAS MINED, PROCESSED THROUGH AN INITIAL CRUSHING OPERATION AND LOADED ONTO RAILROAD CARS. AT THE DAVE JOHNSTON PLANT, THE COAL WAS UNLOADED AND STOCKPILED IN A CONICAL READY PILE, ISOLATED FROM THE OTHER DAVE JOHNSTON OPERATIONS TO PRECLUDED CONTAMINATION. SIMILARLY, THE FEED AND SILO SYSTEMS OF UNIT 1 WERE CLEARED OF DAVE JOHNSTON COAL PRIOR TO INTRODUCTION OF THE JIM BRIDGER COAL.

APPROPRIATE HERE IS A DESCRIPTION OF THE PILOT PRECIPITATOR:

1. OWNERSHIP - SF HAD JUST COMPLETED THE ASSEMBLY OF A NEW PILOT PRECIPITATOR IN CANADA, SIMILAR TO THE UNIT USED FOR AUSTRALIAN TESTS FOR VALES POINT, LIDDELL AND WALLERAWANG POWER STATIONS.

2. SECTIONALIZATION - THREE IN LINE FIELDS, EACH EQUIPPED WITH A HIGH VOLTAGE ELECTRICAL SUPPLY AND RAPPING MECHANISM FOR COLLECTING PLATES AND FOR DISCHARGE ELECTRODES.

3. PHYSICAL SIZE - APPROXIMATELY 31' LONG x 12' WIDE x 20' HIGH, CAPABLE OF TREATING A NOMINAL 2,500 ACFM OF FLUE GAS.

4. GAS FLOW DISTRIBUTION - TWO PERFORATED PLATES PRECEDED THE INLET FIELD. ADDITIONALLY, FIELD ADJUSTMENT OF SEVERAL GUIDE VANES AND BAFFLES ATTACHED TO THE SECOND SCREEN ACCOMPLISHED A UNIFORM GAS DISTRIBUTION ACROSS THE FACE OF THE INLET FIELD.

5. INLET DUCTS - TWO INLET DUCTS, ONE FOR COLD GAS TESTS AND ONE FOR HOT GAS TESTS, WERE DESIGNED SPECIFICALLY TO ACHIEVE ISOKINETIC SAMPLING. DUCTS WERE EQUIPPED WITH A HEAT EXCHANGER USED AS A COOLER (AIR AND WATER) AND ELECTRIC HEATERS TO CONTROL GAS TEMPERATURES. PILOT UNIT CASING WAS FITTED WITH ELECTRIC HEATERS TO MAINTAIN NEAR CONSTANT TEMPERATURES THROUGH THE PRECIPITATOR.

6. ELECTRODES - FOUR PARALLEL GAS PATHS, SPACED 9-3/4" AND FORMED BY COLLECTING ELECTRODES 6'-6'3/4" HIGH, TOTAL COLLECTING SURFACE 826.7 SQ. FT. DISCHARGE WIRES WERE SPIRAL WOUND, ROUND SECTION, 0.1" DIAMETER.

7. RAPPERS - TUMBLING HAMMER TYPE, FOR BOTH COLLECTING AND DISCHARGE ELECTRODES.

8. ELECTRICAL - FULL-WAVE SELENIUM RECTIFIERS, 60KV EACH AT 45 MA RATING.

9. ASH REMOVAL - A CHAIN CONVEYOR DISCHARGES COLLECTED ASH INTO THREE STEEL CANS UNDER EACH COLLECTING HOPPER, THROUGH AN AIR-TIGHT VALVE.

THE ENTIRE PILOT UNIT WAS HOUSED IN A TEMPORARY, QUONSET TYPE BUILDING ADJACENT TO DAVE JOHNSTON UNIT 1. THE BUILDING WAS PROVIDED WITH HEAT, LIGHTS AND WORK BENCHES. OFFICE SPACE AND TELEPHONE SERVICE, INCLUDING A COMPUTER TERMINAL, WERE PROVIDED IN A NEARBY TRAILER. CALCULATIONS TO VERIFY THE VALIDITY OF A TEST RUN WERE QUICKLY MADE PRIOR TO THE COMMENCEMENT OF THE NEXT TEST.

TESTING PROGRAM

FOLLOWING A SERIES OF PRETESTS TO CHECK EQUIPMENT OPERATION AND TO TRAIN PERSONNEL, THE TEST PROGRAM MOVED TO THE FIRST TEST RUN ON NOVEMBER 29, 1971. DURING A THREE-MONTH PERIOD, NOVEMBER, 1971 THROUGH JANUARY, 1972, OVER 175 TEST RUNS WERE MADE, ENDING WITH THE FINAL TEST JANUARY 26, 1973. NOT ALL OF THESE TESTS WERE RUN WHILE FIRING JIM BRIDGER COAL. SEVERAL TEST SERIES WERE UNDERTAKEN WITH DAVE JOHNSTON COAL IN ORDER TO CONFIRM VALID PERFORMANCE OF THE TEST UNIT, TO MAKE CORRECTIONS IN OBSERVED TEST UNIT "SNEAKAGE" AND TO CONSERVE JIM BRIDGER COAL FOR VITAL TESTS. OF THE 175 TEST RUNS, 128 WERE CONDUCTED WHILE UNIT 1 WAS FIRING JIM BRIDGER COAL. ALL TESTS WERE CONSIDERED GERMANE TO THE ISSUE TO THE JIM BRIDGER ELECTROSTATIC PRECIPITATOR DEVELOPMENT PROGRAM.

IN CONDENSED FORM, THE COMPLETE PROGRAM EVALUATED THESE ISSUES:

1. FLUE GAS TEMPERATURE EFFECTS - DEVELOPMENT OF PILOT MIGRATION VELOCITIES.
2. MOISTURE CONDITIONING
3. AMMONIA CONDITIONING
4. VOLT-AMPERE RELATIONSHIPS

5. COAL ANALYSES RELATIONSHIPS

6. DUST ANALYSES RELATIONSHIPS

THE PRIMARY ISSUE TO BE RESOLVED WAS THE DETERMINATION OF THE REQUIRED SPECIFIC COLLECTION AREA TO ACHIEVE THE SPECIFIED 0.02 GRAINS/ACFM EMISSION GUARANTEE FOR THE FULL SCALE JIM BRIDGER ELECTROSTATIC PRECIPITATORS.

SOMETHING NOW ABOUT THE TEST CREW AND ITS MAKE UP OF PERSONNEL. SF-CARBORUNDUM WAS REQUIRED BY CONTRACT TO BE RESPONSIBLE FOR THE CONDUCT OF THE TESTING PROGRAM. PACIFIC POWER, IDAHO POWER, BECHTEL AND OTHER OWNER SELECTED CONSULTANTS WERE INVOLVED.

A TYPICAL TEST CREW WAS MADE UP OF APPROXIMATELY 10 PEOPLE; TWO TO TEST FOR PARTICULATES, ONE TO TAKE ORSTAT READINGS, TWO TO WEIGH AND HANDLE THE DUST DISCHARGE, ONE LABORATORY TECHNICIAN, TWO FOR COAL SAMPLING, ONE FOR CALCULATIONS AND ONE TO ACT AS THE TEST CREW SUPERVISOR. NOT ALL TESTING WAS CONDUCTED ON AN AROUND-THE-CLOCK BASIS, THE EARLY TEST RUNS AND PARTICULARLY THOSE CONDUCTED WHILE FIRING DAVE JOHNSTON COAL WERE UNDERTAKEN WITH ONE SHIFT DURING THE DAY. IN THE LATER STAGES, WITH THE JIM BRIDGER COAL SUPPLY SUBSTANTIALLY REDUCED, A 24-HOUR TESTING EFFORT PREVAILED.

AT THE PEAK INTENSITY OF THIS TESTING PROGRAM, IN EXCESS OF 30 TECHNICAL INDIVIDUALS WERE INVOLVED AT THE SITE. NOT INCLUDED IN THIS COUNT WERE A NUMBER OF MAINTENANCE AND CONSTRUCTION INDIVIDUALS. A SPECIAL REPRESENTATIVE OF THE PLANT OPERATING STAFF MAINTAINED APPROPRIATE LIASON AND COORDINATION WITH DAILY PLANT OPERATIONS WHICH BY NECESSITY COMMANDED THE HIGHEST PRIORITY.

It has been estimated that the entire test program cost in excess of $600,000 including the supply and shipment of the Jim Bridger test coal.

PILOT TEST CORRELATION TECHNIQUE

The SF technique for sizing an electrostatic precipitator from pilot test data is unique, was required to be specified in advance of the pilot testing effort and perhaps controversial in some aspects. This presentation will not discuss in depth the numerical or theoretical details of the SF approach to precipitator sizing but will describe the basic elements as applied to the Jim Bridger precipitator development program.

For Jim Bridger, pilot unit to full-size precipitator correlations were made by application of a modified form of the conventional Deutsch collection efficiency formula.

$$\mathrm{Eff} = 1 - e^{-\left(\frac{A}{Q} W_K\right)^K}$$

where: Eff = collection efficiency
A = collecting surface area
Q = flue gas volume flow rate
W_K = migration velocity, corrected
K = correction factor
e = base of natural logarithms

The unit K is a correction factor, developed by experience and applied to adjust the calculation for theoretical vs actual performance differences. W_K is an effective migration velocity developed from migration velocities measured in the pilot test work and corrected by a number of factors evaluating the influence of dust, gas and physical properties on full-size pre-

CIPITATOR PERFORMANCE.

MIGRATION VELOCITY CORRECTION FACTORS UTILIZED IN THE JIM BRIDGER CALCULATION WERE:

1. TEMPERATURE CORRECTION - TO DEVELOP A CORRELATION BETWEEN FLUE GAS TEMPERATURE AND REQUIRED COLLECTING AREA.

2. GEOMETRIC CORRECTION - LEAKAGE AND SNEAKAGE AREA IN THE PILOT UNIT IS LARGE IN COMPARISON TO SIMILAR BOUNDARY AREAS OF THE FULL-SIZE UNIT REQUIRING A CORRECTION FACTOR TO COMPENSATE FOR DIFFERENCES. THIS FACTOR DEVELOPED FROM EXPERIENCE.

3. UNCERTAINTY CORRECTION - A SAFETY FACTOR DEVELOPED BY EXPERIENCE.

4. DUST CONCENTRATION CORRECTION - A FUNCTION OF DUST CONCENTRATION, PILOT UNIT VS FULL-SIZE UNIT. AGAIN, A RELATIONSHIP DEVELOPED BY EXPERIENCE.

5. MOISTURE CORRECTION - A FUNCTION OF PILOT UNIT FLUE GAS MOISTURE VS FULL-SIZE UNIT GAS MOISTURE, RELATIONSHIP DEVELOPED IN THE PILOT TEST WORK.

PARTICLE MIGRATION VELOCITY IS DEPENDENT ON CURRENT DENSITY WITHIN THE PRECIPITATOR. TO LIMIT LARGE CORRECTIONS IN MIGRATION VELOCITY, SF ELECTS TO OPERATE THE PILOT UNIT AT THE SAME CURRENT DENSITY AS WILL BE THE CURRENT DENSITY OF THE FULL-SIZE UNIT. FROM EXPERIENCE, A CURRENT DENSITY FOR JIM BRIDGER WAS ESTABLISHED AS 14.5 MICRO AMPERES PER SQUARE FOOT OF COLLECTING SURFACE. PERTINENT PILOT TESTS WERE CONDUCTED AT THIS SAME CURRENT DENSITY.

PILOT TEST RESULTS

A CONDENSATION OF THE RESULTS OF THE PILOT TEST WORK IS IN ORDER. THE PRINCIPAL EFFORT IS TO DESCRIBE THE SELECTION OF THE FINAL SIZE FOR THE FULL-SCALE JIM BRIDGER PRECIPITATORS.

1. JIM BRIDGER PRECIPITATOR SIZE - THE ORIGINAL INTENT WAS TO SET THE DAVE JOHNSTON UNIT 1 PULVERIZER COAL GRIND TO DUPLICATE THE JIM BRIDGER SPECIFIED GRIND OF 70% THROUGH A 200 MESH SCREEN. AS THE PROGRAM PROGRESSED, THIS TECHNIQUE PROVED NOT ONLY DIFFICULT TO MAINTAIN, BUT THERE WAS EVIDENCE THAT COAL GRIND SIGNIFICANTLY INFLUENCED PILOT PRECIPITATOR PERFORMANCE. AS THE COAL GRIND BECAME MORE COARSE, PILOT PRECIPITATOR PERFORMANCE INCREASED.

   EVEN AFTER A CAREFUL ANALYSIS OF ALL TEST DATA THE ISSUE OF COAL GRIND VS ASH PARTICLE SIZE WAS NOT CLEARLY DEFINED. HOWEVER, THERE WAS CLEAR EVIDENCE THAT CORRELATION OF A VERY FINE COAL GRIND WOULD PREDICT A VERY LARGE FULL-SIZE JIM BRIDGER PRECIPITATOR. CORRELATION OF THOSE TESTS WITH A COARSER COAL GRIND, APPROXIMATING THE JIM BRIDGER SPECIFIED 70% THROUGH A 200 MESH SCREEN, SUGGESTED A PRECIPITATOR SIZED WITH A SPECIFIC COLLECTING AREA OF 375 (375 SQ. FT./1000 ACFM).

   THE SERIES OF PILOT TEST RUNS IDENTIFIED WITH THE EXTREMELY FINE COAL PULVERIZER GRIND WAS DESIGNATED AS SERIES A. THOSE TEST RUNS CONDUCTED WITH A PULVERIZER GRIND APPROXIMATING THE 70% THROUGH A 200 MESH SCREEN (JIM BRIDGER SPECIFICATION) WERE DESIGNATED as Series C. A Jim Bridger precipitator sized on Series A data would equate to an SCA of 586. As previously mentioned, a Jim Bridger precipitator sized on the Series C tests would equate to an SCA of 375.

SF developed their recommendation based on this rational:

a) A size based on Series C tests is the most representative of full-scale Jim Bridger circumstance.

b) It is possible, but not likely, that conditions supporting the fine grind, Series A tests could prevail at some time in the operation of the Jim Bridger plant.

c) A reserve, empty collecting space of 20% has been specified in the design.

d) Allocation of this 20% reserve space to performance protection should the extreme circumstances of Series A conditions prevail, suggests that the pruduent full-size design be 586 SCA (Series A, fine grind requirement) less the 20% reserve space margin.

e) This calculation results in a recommended Jim Bridger precipitator size of 488 SCA.

f) The size of 488 SCA includes a margin sufficient to compensate for the "band" of results that would be the complete Series C tests (tests representing the most likely Jim Bridger circumstances).

The Owners and their Engineers concluded that this was a conservative, though not extravagant conclusion and the Jim Bridger Unit 1, 2 and 3 precipitators were cast into final manufacture based on a size equal to 488 sq. ft. of collecting surface for each 1,000 acfm.

2. Moisture Conditioning - Results indicated practically no susceptibility to moisture conditioning.

3. Ammonia Conditioning - Results suggested a very small increase in precipitator performance with ammonia conditioning is possible.

4. High Temperature Results - The test work evaluating high temperature effects was limited. This data, plus data from Australian test work did not suggest an advantage to consideration of a "hot" precipitator design.

FIRST PERFORMANCE TESTS

UNIT 1 WAS THE FIRST TO UNDERGO TESTING FOR PERFORMANCE. TEST A WAS CONDUCTED DURING THE PERIOD OF APRIL 14-30, 1975. UNIT 1 HAD BEEN IN OPERATION IN EXCESS OF 4,000 HOURS WITH AT LEAST THE REQUIRED MINIMUM OF 2,000 HOURS WITHOUT ADJUSTMENT OR MODIFICATION. TESTING PRODUCED COLLECTION EFFICIENCIES RANGING FROM 99.5 TO 99.6% (GUARANTEE - 99.33%) WITH ALL GAS PATHS IN SERVICE. WITH ONE GAS PATH OUT OF SERVICE EFFICIENCIES OF 99.4% WERE RECORDED, WELL WITHIN THE SPECIFIED 98.73% UNDER THIS CONDITION. PARTICULATE EMISSIONS IN EVERY CASE WERE WELL WITHIN THE GUARANTEED 0.02 GRAINS/ACF.

TEST B FOR UNIT 1 AND TEST A FOR UNIT 2 WERE SCHEDULED FOR FEBRUARY, 1976 AND MARCH, 1976. IN FEBRUARY, 1976, PRELIMINARY TESTS OF UNIT 1 INDICATED PERFORMANCE OF QUESTIONABLE QUALITY. AS A RESULT, FURTHER OFFICIAL PERFORMANCE TESTS OF UNIT 1 AND 2 WERE DEFERRED PENDING TECHNICAL INVESTIGATION TO DETERMINE THE CAUSE OF THE REDUCED PRECIPITATOR CAPABILITY. THESE INVESTIGATIONS ARE CONTINUING AS OF THIS DATE.

BASED ON EARLIER EXPERIENCE, IT WAS SUSPECTED THAT THE POOR PRECIPITATOR PERFORMANCE, INDICATED BY UNSATISFACTORY ELECTRICAL READINGS AND BACK CORONA, WAS CAUSED BY PHENOMENA LINKED TO ELECTRODE DEPOSITS FROM OIL FIRING DURING UNIT START-UP. SATISFACTORY PERFORMANCE WOULD RETURN WITHIN A FEW DAYS FOLLOWING START-UP.

CORRECTIVE AND INVESTIGATIVE MEASURES NOW ACTIVE FOR SOLUTION OF THIS PROBLEM ARE:

A) Removal of a representative sample of electrodes from Unit 1 and Unit 2 and subject them to laboratory examination to determine the nature of the coating.

B) Clean emitting electrodes in one electrical section, clean collecting electrodes in a second electrical section, and clean both emitting and collecting electrodes in a third section. Observe electrical and operating characteristics of each section before and after cleaning.

C) An improved operating procedure has been issued to cover unit start-up, shutdown and low-load operation.

SF-Carborundum are confident that a complete and lasting solution will be developed. As of this very date, Unit 1 and Unit 2 are available for full load operation without precipitator imposed limitations.

CONCLUSION

Pacific Power, Idaho Power, their Engineers and SF-Carborundum consider the engineering effort of the Jim Bridger Electrostatic Precipitator development program to be the most thorough and technically sound of any United States precipitator installation. Hundreds of individuals have been involved over a period in excess of seven years. The Jim Bridger electrostatic precipitators prove that high efficiency collection of low sulfur, Western coal ash is possible and practical, "cold side."

ACKNOWLEDGEMENTS

Many individuals from several organizations contributed significantly to the success of the Jim Bridger Electrostatic Precipitator Development Program. The author has chosen to list, for purposes of personal recognition, the names of most of those responsible for this major engineering accomplishment. Omissions there are. For this human deficiency the author apologizes. There has been no conscious intent to omit appropriate recognition.

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- Kenneth Watson

OPERATING EXPERIENCE WITH ESP CONDITIONING IN RELATION TO AN ELECTROSTATIC PRECIPITATOR UPGRADING PROGRAM

Scott H. Cragle
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OPERATING EXPERIENCE WITH ESP CONDITIONING IN RELATION TO AN ELECTROSTATIC PRECIPITATOR UPGRADING PROGRAM

SCOTT H. CRAGLE, PENNSYLVANIA POWER & LIGHT CO.

Abstract

This report summarizes the major areas of an ongoing electrostatic precipitator (ESP) upgrading program at Pennsylvania Power and Light Co. Particular emphasis is placed on the area of flue gas conditioning including eighteen months of operating experience with Apollo Chemical Corp. products, a planned SO_3 injection trial, and other potential agents.

Results of ongoing trials of Apollo products at Montour SES on two 750 MW bituminous coal fired units are presented. Performance data, operating problems and cost information are covered. Eight chemical formulations have been injected into various combinations of four locations in efforts to improve product performance, solve air heater plugging problems and improve distribution of the product on the fly ash. A significant reduction of emissions on a difficult fly ash has been noted although only half as great a reduction as originally expected based on tests run with a good collecting fly ash. Air heater plugging problems have been eliminated.

The report also covers some of the practical considerations and potential problems that can occur in evaluating proprietary additives.

Also, the report briefly summarizes projects on six of the Company's ESP's including work in the following areas: fuel quality effects; flue gas conditioning, rappers; voltage controls; sectionalizing; gas flow distribution; operating and maintenance and additional dust collector capacity.

May 11, 1976

Pennsylvania Power & Light Company (PL) began a series of efforts to upgrade its existing electrostatic precipitator (ESP) capacity in mid-1973. At that time it had become apparent that the particulate emissions from the two new boilers at PL's Montour station did not meet consistently acceptable levels. The Montour station has two 750 Mw units with pulverized bituminous coal boilers. The ESP's were built in a chevron configuration and designed for a 99.4% efficiency for coals with a 1.5% to 3.0% sulfur and an efficiency of 99% for coals with a 1.0% to 1.5% sulfur content. More detailed information is shown in Table 1.

Initial Performance Tests (9/73-11/73)

An initial series of efficiency tests on Unit 1 resulted in efficiencies from 90.9% to 95.4%. A compliance test by the Pennsylvania Department of Environmental Resources showed emissions 9 to 10 times the emission limit of 0.1 lb./10^6 BTU. Another series of outlet tests was run to determine the effect of the sulfur content of the different fuels burned at the plant and also to determine the feasibility of using the two recently installed Lear Siegler opacity monitors as indicators of ESP performance.

The tests showed a relation between optical density and dust flow from the precipitator (Figure 1). As a result of this test, the monitors have become an integral tool in monitoring precipitator performance. The tests showed that the ESP's performed within regulation with certain high sulfur fuels, but performed poorly with the low sulfur fuel from our Greenwich Mine, a major fuel source for the plant. The test points are numbered on Figure 1 in the order that the tests were performed. Tests 1 through 4 show that we didn't observe the full effect of a low sulfur coal until it had been burned for several consecutive days. This was due to the residual fly ash in the ESP. Tests 5 through 8 indicate that the good effects from a high sulfur coal were not noticed on the first day after burning a low sulfur coal, but only on the second and third day. Test 8 was a lower sulfur coal, but high levels of SO_3 were reported during the test, possibly due to a high vanadium content in the ash.

EXPERIENCE WITH THE APOLLO FLY ASH CONDITIONING AGENT

Based on the test results, we viewed flue gas treatment to be the major solution to the ESP performance problem at Montour. Ongoing work done in other areas is covered briefly later in the paper. After a short study (1/74) of conditioning alternatives, we chose Apollo (2/74) on the basis of a low first cost and a short three month lead time for installation. We thought that Apollo would be a quick, low cost demonstration of the effectiveness of fly ash treatment on our problem fuels. Other alternatives required large capital investments and nearly a one year lead time to set up any demonstration.

Our initial plans were for the trial to last three months on one 750 MW unit. The cost projection was $150,000 (20¢ Kw) to install the system and $90,000 for three months of chemical supply. The system has now been expanded to both Montour units for a cumulative capital cost of about $400,000 ($200,000 per unit or 27¢/Kw). The cost of chemicals to treat both units is $650,000 per year (0.0067¢/KW HR).

Performance LPA-40® (9/74-1/75)

All conclusions about ESP performance in this report are based on data taken by PL personnel.

The Apollo product used in the initial testing was LPA-40. This was a sulfamic acid (PL analysis) based multicomponent product that is injected into the flue gas upstream of the boiler economizer. Through breakdown to NH_3 and SO_3 it is intended to affect flyash resistivity, agglomeration, and space charge in the flue gas.

Because of a 3 month strike at PL, initial operation of the Apollo system was delayed until September, 1974. We quickly discovered that we were limited to injecting less than 0.1 Gal/Ton of coal (25 GPH) because of increased pressure drop across the Lungstrom air heaters. A series of emission tests were run at this feed rate, and two opacity monitors made continuous observations. Table 2 shows the improvement that occurred with this Apollo additive using our regulated emission limit (0.1 LB/10^6 BTU) as a base line (1x). Emissions were reduced by 30-45%.

Internal inspection of the air heater showed no adverse effects related to the high pressure drop. Whenever treatment was shut off for a few days the pressure drop would return to normal. Most likely, the plugging was due to unvaporized product on the air heater or recombination of vaporized product to ammonium bisulfate.

During the next phase of the test program, our efforts were centered on reducing the air heater plugging to allow higher treatment rates. These five approaches were taken (in chronological order):

1. Product Atomization (10/74) - air atomized nozzles were added to the system to assist in faster breakdown of the product. No major improvement resulted from this change.

2. Flow Distribution Tests (11/74) - We suspected uneven treatment of the flue gases might be a significant problem because of the location of the injection ports (Figure 3, Phase I). Tests run at the inlet of the air heater showed evidence that 80% of the gas stream was getting uniform treatment and 20% of the gas stream was receiving lower than normal concentrations of the product. We are not certain if this situation improves upstream of the precipitator. Even if it is a significant problem, no ready solution is available to us to allow more even treatment of the flue gas.

3. LPA - 40, Neutral Formulation (11/74) - Apollo changed the major component of their product from sulfamic acid to ammonium sulfate (results of PL analysis) indicating that the new product decomposed faster. No significant improvement in performance resulted from the change of product.

4. Flue Gas Temperature at the Point of Injection (1/75-4/75) - Flue gas temperature at the point of injection is apparently critical to the application of LPA - 40. Injecting the product into the flue gas stream at too high a temperature results in the chemical combination of the products with the fly ash. Also, production of SO_3 at high temperatures is not useful because the SO_3 will convert to SO_2 by the nature of the SO_2 - SO_3 chemical equilibrium. Injecting at too low a temperature, the product does not decompose as completely and as rapidly as necessary and air heater plugging results. Our problem was low temperature at the injection point.

   At the start of the program, Apollo indicated that the ideal temperature of injection was 800-1000° F. Following their change to the neutral formulation, Apollo indicated that they had seen good results at injection temperatures above 1200° F and that temperatures as high as 1500° F would be optimum for their product. Currently they believe the optimum temperature is 1250-1300° F.

   Figure 3, a boiler cross section, shows the locations we have injected LPA-40 with the associated flue gas temperatures for both full boiler load and half boiler load. Note how the temperature varies with load.

   Phase I (950-750° F) - This is the original penthouse location where we have had the best results so far. Injection here at the rate of 0.1 gal/ton resulted in air heater plugging.

   Phase II (2300-2200° F) - Injection above the firing zone at high rates (0.4 gal/ton) had no effect. No effect was seen on precipitator performance and no air heater plugging occurred.

   Phase III (1650-1250° F, upper ports; 1750-1350° F, lower ports) - Injection into the side of the boiler into the superheat area showed no effect at full load. No effect was observed on precipitator performance and no air heater plugging occurred.

   Phase IV - The problem of finding an injection point with the right temperature caused the next phase of testing. Product was injected through the Phase I location during periods of high boiler load and injected into the Phase III location during periods of low boiler load in an attempt to keep the injection temperatures acceptable. This approach was intended to reduce the chance of air heater plugging while providing continuous treatment of the fly ash at all boiler loads. No improvement in precipitator performance or in the plugging problem resulted from this approach.

5. Revised Apollo Chemical Formulations (5/75 - Present)

   LPA 401, Catalytic Formulation - This is the LPA-40 formulation with a metal added to cause faster breakdown. This formulation caused more air heater plugging problems than the normal LPA 40.

LPA 402, Agglomerator Formulation - This formulation contains an organic compound in trace quantities. Results with this have been good. Air heater plugging has been eliminated and ESP performance has been equivalent to LPA-40.

LPA 403, Combination of LPA 401 and 402 - No trial was run due to the problems with 401.

LPA 404, Non-Ammoniated formulation SO_3 - release agent - This was designed to produce SO_3 only. Results with this chemical showed no improvements in SO_3 concentration in the flue gas or improvement of the in-situ resistivity of the fly ash.

LPA 405, Another Non-Ammoniated formulation - This is similar to 404, but has not been tried due to handling concerns resulting from its low pH of 1.0.

LPA 402A, Agglomerator Formulation - this is similar to 402 only with a different organic compound. It is more volatile than 402. Results with this chemical seem identical to those with 402 and it is the formulation presently being used.

PERFORMANCE - LPA 402 AND 402A

Table 3 shows results for Unit 2 with LPA 402A in the same format as Table 2 shows results for LPA 40 on Unit 1. The Unit 2 precipitator in general performs better than Unit 1 due to some variations in physical flow arrangement. The results indicate that 402 and 402A are similar performers to LPA 40 in the Montour boilers when injected at the penthouse location.

During April and May, 1976, a series of ESP efficiency tests are being run on Montour 2. The tests are being run with high sulfur coal, low sulfur coal, and low sulfur coal conditioned with LPA-402A.

Resistivity

Figure 4 plots in-situ resistivity taken by Apollo and Southern Research Institute using a point to plane probe. These were taken at various times during our test program. The scatter of the data is evident in this graph and, as a result, it is difficult to make conclusive statements about it. By looking at selected points, we have noticed certain trends, however. Increased sulfur content of the fuel has the expected effect of lowering the resistivity. Lower temperatures seem to produce lower resistivities as expected. The school of thought at PL, which has not been fully investigated, is that lower flue gas temperatures may help the performance of the Apollo flue gas treatment by lowering the resistivity. In general, for a low sulfur fuel, it seems that the Apollo LPA 402 treatment lowers the resistivity about one half of an order of magnitude from the low 10^{11} OHM-CM to the high 10^{10} OHM-CM.

At this point our data on available SO_3 in the flue gas is incomplete. Apollo picks up most of the sulfur compounds in hopper samples of fly ash.

It is difficult to draw conclusions from either resistivity analysis or SO_3 analysis because of the variable sulfur content of fuel burned at Montour. Monitoring minute-to-minute changes in the sulfur content of the coal is difficult.

General Observations

These are some suggestions that will be helpful to a company considering a test program with a proprietary chemical flue gas conditioning additive.

1. PL has found that trials planned to be quick and inexpensive can turn out to be more involved than initially anticipated. If you are using expediency for justification of a choice of a certain alternative, be certain that your trial program will be expedient.

2. Prior to the start of a test program, define the ground rules for the evaluation program.

    a. What is the expected mechanism or action of the additive?

    b. What parameters will be used to measure the performance of the additive (resistivity, SO_3 concentration, NH_3 concentration, fly ash analysis, etc.)? Ideally you will set up a program to measure the additives effect in more ways than just precipitator performance. In this way you can determine if changes in ESP operation are due to the additive or other parameters. Also it will be easier to troubleshoot any lack of performance.

3. Get a committment from the vendor prior to the program as to the "ideal" temperature and location for injection of the product. If possible, investigate the temperature (and its variation with boiler load) at proposed injection locations before installing ports and starting the trial.

4. Be sure your injection system is capable of adequate distribution of the additive in the flue gas. Cutting corners in this area may well limit the success of your trial.

5. Use great care in installing the injection system and, if possible, observe its operation (injecting water) from inside the boiler during an outage. An unfortunate error in our installation resulted in 16 economizer leaks and a six day unit outage.

6. To demonstrate the capabilities of the ESP, arrange to burn some higher sulfur fuels and test the units under ideal resistivity conditions.

7. If your system is installed during an annual outage, allow the boiler to become dirty before testing emissions and drawing conclusions on the performance of an additive product. PL's Chem Lab feels that a clean boiler after an outage acts as a catalytic bed in the conversion of SO_2 to SO_3. During the period following an outage the higher levels of SO_3 result in somewhat better precipitator performance. Also the precipitator is usually cleaner following an outage.

   Also, note any modifications that are made during an outage to the boiler such as addition of economizer of other tube surface to the boiler. These may affect the amount of SO_3 formed by catalytic action or they may affect the temperature (thus resistivity) of the fly ash.

8. Often, ESP performance improvements attributed to additives are due, in part, to changes in other parameters. Data in this report was taken by PL under carefully controlled conditions to minimize this problem.

   a. During both baseline and additive testing, maintain uniform operating conditions in the boiler (coal quality, excess air, gas temperature, mill classifier setting, etc.). Variations in these parameters can cloud your results.

      Continuous reading SO_2 and opacity monitors have been a useful tool in quickly determining relative changes in these parameters. Absolute values from these instruments depend on monitor reliability and maintenance.

   b. Familiarize yourself with the operation of the ESP. Use consistent set points for rappers and automatic voltage controls (at sparking threshold) during all testing. Observe the power levels during emission testing to detect any abnormal conditions that may invalidate results of that particular test.

9. At Montour we have found that changes in the type of coal burned in the boiler do not have the full effect on the precipitator for a number of days. Be careful not to draw positive or negative conclusions too soon after adjusting any operating parameter.

10. By following these guidelines and being open-minded towards alternate chemical formulations and injection locations, recommended by the supplier, PL has benefited with improved performance. Try to identify all possible approaches before the start of the trial.

Summary

The original plan for the Apollo trial was a quick, simple, and inexpensive demonstration with hopes for impressive results. Although the trial turned out to be long, involved, and costly with less than spectacular results, day to day emissions at the Montour plant have been reduced significantly through the use of Apollo.

PL'S PARTICULATE COMPLIANCE PROGRAM (1/75-Present)

PL's particulate compliance program is a series of trials in 8 investigative areas on six of PL's precipitators.

Starting with the formation of a 5 man Air Quality Project Team in January, 1975, PL started a two year investigative program to bring its units into full day-to-day compliance with Pennsylvania air quality regulations. In the past, PL had attempted to have ESP's comply by making modifications that seemed most promising. Often, too many changes were made to a unit at one time. This made it difficult to determine what, if any, improvement was gained, and if the changes could be applied to other units with similar problems. Previous work done on Montour No. 1 is a prime example. This work included:

A. Modification of rapper shafts and the rapping program.

B. Adjustment of electrode control set points.

C. Trials with updated high voltage controls on sections of the precipitator.

D. Installation of turning vanes in an attempt to improve gas flow distribution.

E. Experimentation with adding weight to the electrode wire tensioning weights.

F. Collecting plate realignment and wire support frame stabilization.

G. Blending of high sulfur coal with low sulfur coal at the mine in order to raise the overall sulfur level of the coal.

H. Adjustments to operating parameters such as excess oxygen and flue gas temperature.

I. Apollo fly ash conditioning trial.

The combination of concurrent changes made it extremely difficult, if not impossible, to reasonably evaluate the effect of any single change. The intent of the present program is to produce as conclusive a result as possible about the effects of various modifications that will be made. Table 1 lists the design parameters of the six precipitators.

This is a brief summary of the work in each investigative area.

1. Fuel Quality Effect

   The root of the problem with precipitators is often the type of fuel burned. PL uses a wide variety of coals ranging from 10% to 30% ash and from 0.8% to 2.5% sulfur. A study is being done on all of the six units to determine the effects of the various fuels on the dust collector performance. At the end of the study PL will consider the economics and practicality of achieving compliance by controlling the quality of fuel shipped to our plants. This approach is possible by using the capabilities of cleaning plants at our mines and coal blending.

2. Flue Gas Conditioning

   A. Apollo - Use of the Apollo product will continue unless a more effective gas conditioning system is demonstrated or additional collecting capacity is installed at Montour. PL has an agreement with Pennsylvania regulatory authorities to continue any reasonable activity that lowers emissions at Montour.

   B. SO_3 injection, Brunner Island Unit 1 - a Wahlco SO_3 injection utilizing a liquid SO_2 vaporizer system will go into trial service on this unit during August, 1976. A series of tests with the SO_3 injection will be run to get data similar to that in Tables 2 and 3. With this data, we hope to assess how effective such a system would be at Montour in comparison to the Apollo system. The total capital cost of the trial system on this 350 Mw boiler is $445,000 ($1.28/Kw) and the expected cost for liquid SO_2 during the six month trial period will be about $70,000 (.0058¢/KwHr). Any permanent installation of this system will include a sulfur burner resulting in $400,000 additional capital cost and a reduction of chemical cost by a factor of 5 or 6. We expect to have results available from this trial by October 1, 1976.

   C. Trimex, Martins Creek 2 - PL's Research and Development group is sponsoring this trial in an effort to demonstrate the feasibility of reducing SO_2 emissions from the boiler. An offshoot of this trial may be conditioning of the fly ash. After almost five months of testing our Chem Lab reports that the Trimex chemical has shown no affect on the SO_2 emissions or on the performance of the ESP's.

D. Other Additives - Although no additional trials are planned, PL is investigating the use of triethylamine, concentrated ammoniacal liquor, Betz proprietary additives, and Nalco proprietary additives. A trial with Nalco's experimental catalytic additive was cancelled after we had unsuccessful laboratory trials.

3. Rappers

Internal inspection of many of PL's precipitators equipped with vibrating rappers have shown heavy dust buildups on the plates. On one of the two Montour Unit 2 precipitators, the vibrating rappers have been replaced with impact rappers. The new rappers' performance is being evaluated.

4. Voltage Controls

Because of the wide variety of fuels burned at PL's plants, we have found that the automatic voltage controls on many of the units are not capable of following the varying conditions in the EPS's. Eight thyristor controls manufactured by L.L. Little are being installed on one of the two precipitators at Brunner Island Unit 3. We hope to evaluate any improvement in the efficiency of the precipitator, any reduction in burnout of wires, and the general ability of the control to follow changes in operating conditions.

5. Sectionalizing

The precipitators at Martins Creek have a relatively high degree of sectionalization and offer a certain amount of flexibility in allowing quick modification of their sectionalizing arrangement. Experiments are being run with varying amounts of plate area per TR set to determine what possible benefit may be gained by increasing the sectionalization of the precipitators on other PL units and future units. This work is in progress at present.

6. Gas Flow Distribution

The Brunner Island 3 precipitator and Montour 1 and 2 precipitators are built with a chevron inlet configuration. There is evidence that the flow distribution is not uniform over the face of the precipitator. Field tests have been run to determine the flow distribution and a model study is being conducted to determine if improvement is possible. If modifications are made, efficiency tests before and after the modifications will be run to determine the effect of improvement in flow distribution.

7. Operating and Maintenance

A complete review of operating and maintenance procedures is being conducted throughout the PL system. Since the precipitators are not a megawatt producing piece of equipment, historically they have received less attention than other equipment. We are reviewing the impact of operating parameters and maintenance on the performance of the precipitators with our plant personnel.

8. Additional Collector Capacity

At the conclusion in early 1977, should the efforts in the seven previously listed investigational areas fail to bring Brunner Island 1 and 3 and Montour 1 and 2 into compliance (Martins Creek 1 and 2 are in compliance) with air quality regulation, additional collector capacity will be installed. The ground work in this area is being done at present to allow for the shortest lead time in purchasing and constructing any additional capacity that may be necessary. The alternatives that are being studied are hot precipitators, cold precipitators, fabric filters, and particulate scrubbers.

Scott Cragle, Project Engineer
Air Quality Project Team
Pennsylvania Power & Light Co.

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TABLE #1 DESIGN DATA ON PRECIPITATORS INCLUDED IN PL'S PARTICULATE COMPLIANCE PROGRAM

Unit	Rating		Flow	SCA	Face Velocity	Treatment Time	Typical	Fuels
	MW		ACFM x 10^3	$\frac{Ft^2}{10^3 \text{ ACFM}}$	$\frac{Ft}{Sec}$	Sec	%S	%Ash
Brunner Island #1	350	Research Cottrell (1961)	550	180	4.2	4.5	1.5-2.5	12-25
		Buell (1965)	550	180	5.2	3.5	1.5-2.5	12-25
Brunner Island #3	750	Western (1969)	2,300	204	5.3	4.6	1.5-2.5	12-25
Montour #1	750	Western (1972)	2,300	204	5.3	4.6	0.8-2.5	12-30
Montour #2	750	Western (1973)	2,300	204	5.3	4.6	0.8-2.5	12-30
Martins Creek #1	160	Buell (1968)	600	274	4.3	6.2	2.0-2.5	12-30
Martins Creek #2	160	Buell (1968)	600	274	4.3	6.2	2.0-2.5	12-30

TABLE #2

MONTOUR

LPA-40 PERFORMANCE

Fuel	Ash	Sulfur	LPA-40	Emissions
Greenwich	18%	1.0%	----	6x - 9x
Greenwich	18%	1.0%	0.1 gal/ton	4x - 5x
Oneida	18%	2.5%	----	1x - 2x
Oneida	14%	2.5%	----	1x (Base)

TABLE #3

MONTOUR #2

LPA-402A PERFORMANCE

Fuel	Ash	Sulfur	LPA-402A	Emissions
Greenwich	20%	1.3%	-----	4x - 12x
Greenwich			0.1 gal/ton	1.5x - 2x
Oneida	12%	2.4%	-----	0.6x

FIGURE I

OPTICAL DENSITY VS. GRAIN LOADING

MONTOUR UNIT 1, NOV. 1973

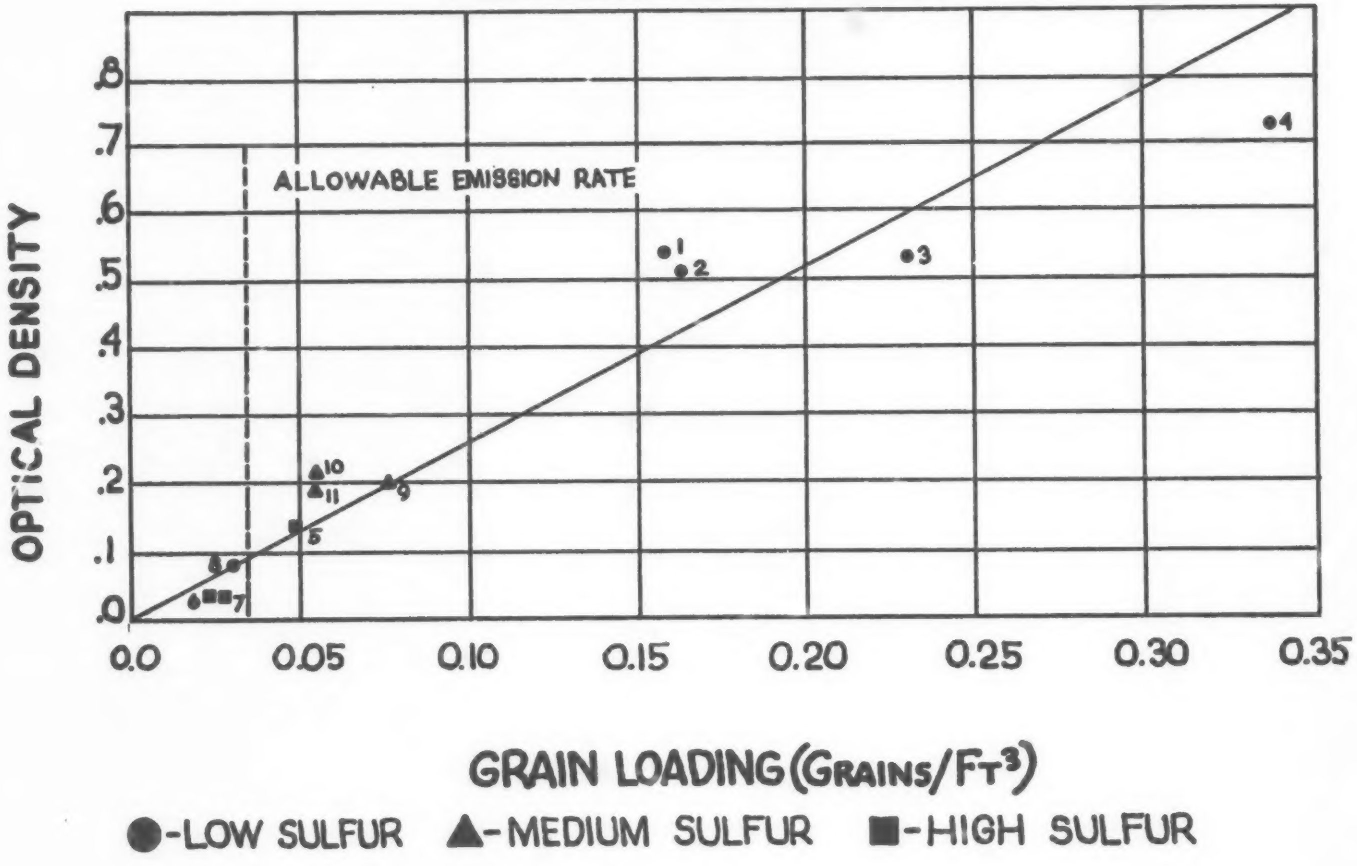


FIGURE 2.
APOLLO INJECTION LOCATIONS
(TEMP. FULL LOAD - TEMP. HALF LOAD)

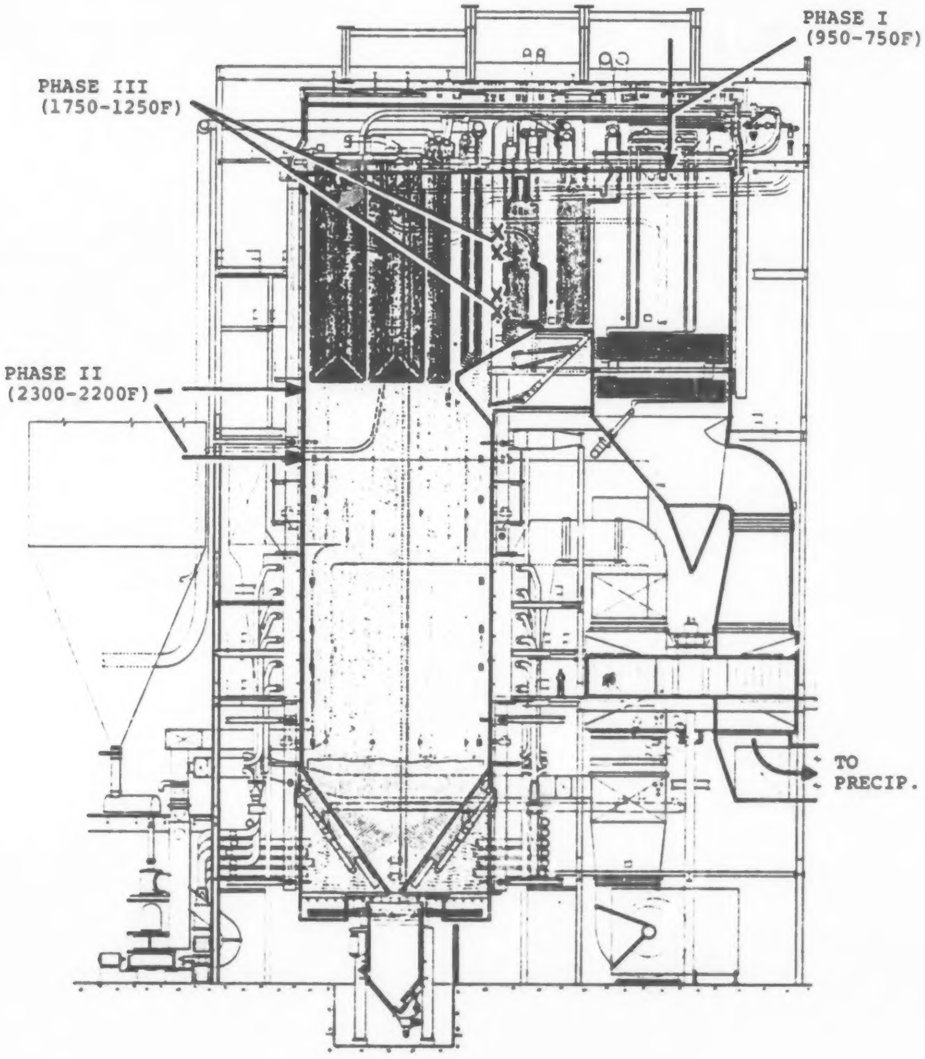
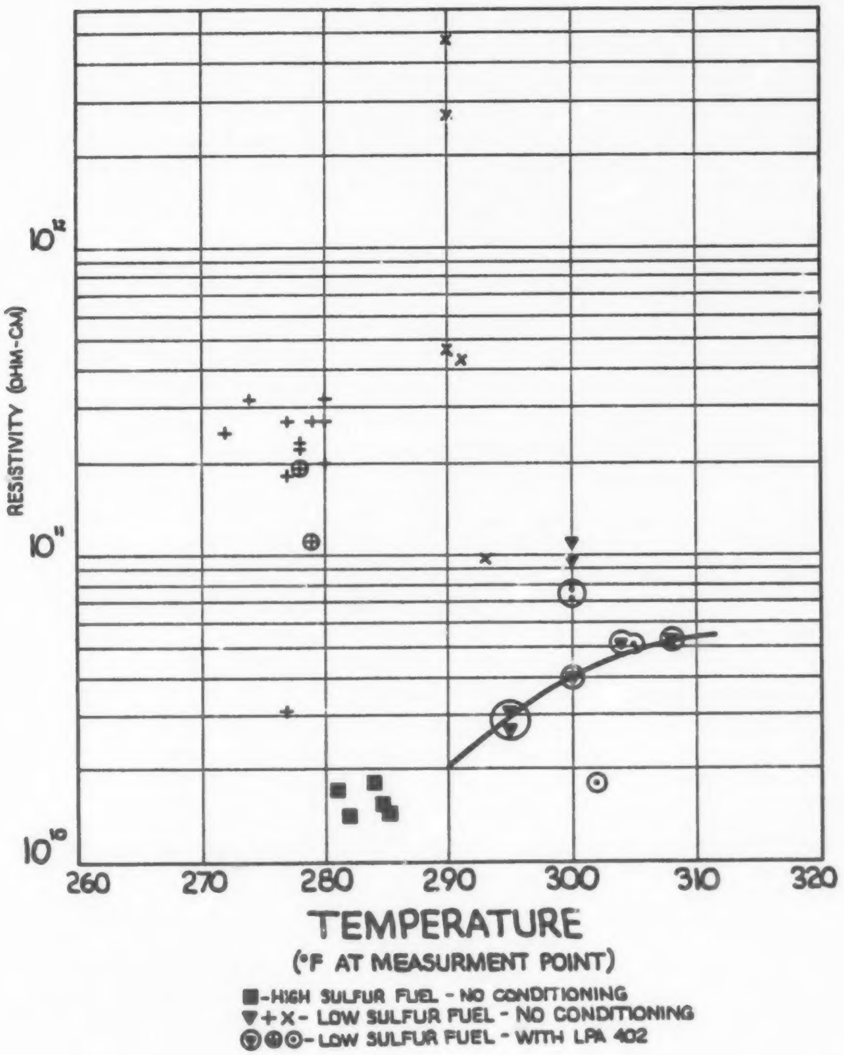


FIGURE 3.

SELECTED RESISTIVITY DATA

WITH AND WITHOUT LPA-402 CONDITIONING



AUSTRALIAN EXPERIENCE WITH FLUE GAS CONDITIONING

Kenneth S. Watson
New South Wales, Australia

SUMMARY

Australian black coal fired in power stations usually contains less than 1% sulphur and has 15 to 30% ash. Target emission levels for new plant are often better than 0.1 gms/m^3 at NTP requiring collection efficiencies of 99.5%. Highly resistive fly ash predicates specific collecting areas in excess of 80 m^2/m^3/s (400 ft^2/1000 cfm) for cold precipitators. Early precipitators frequently had specific collecting areas of less than 50 m^2/m^3/s and performed badly. Gas conditioning has been regularly used to improve collection efficiencies since the middle 60's and to date has been provided on 26 precipitators on boilers from 25 to 500 MW capacity and totalling over 4,000 MW.

Steam and/or water temperature modification, P_2O_5, NH_3 and SO_3 were used experimentally from 1955 onwards. Sulphuric acid was first used operationally in 1964 and anhydrous ammonia or an ammoniacal steelworks waste product have been used since 1965. Triethylamine was first used in 1975 and is currently being evaluated.

The effect of reagents varies from station to station and is also dependent on precipitator type and reagent concentration. Considerable care must be given to the injection systems to allow proper mixing and sufficient residence time. Improvements of migration velocity in excess of 300% have been attained with SO_3/H_2SO_4 and in excess of 200% with ammonia, which is the preferred reagent because of relative safety and convenience. Triethylamine appears to be even more effective than SO_3.

Comparatively few operational problems have been reported and most can be traced to over-conditioning or to poor distribution. Capital costs range from 0.05 to 0.12 \$A/KW and operating costs from 0.01 to 0.05 mils/KW-hr depending on the reagent and injection concentration.

AUSTRALIAN EXPERIENCE WITH FLUE GAS CONDITIONING

Kenneth S. Watson *

1.0 Introduction

Australia is approximately the same area as the continental United States and comprises six States and the Northern Territory (Figure 1). Of the total population of 13 million, 90% occupy the south eastern corner of the continent and the island State of Tasmania. In all States (except Tasmania) coal fired steam power stations provide the majority of electrical energy and will continue to do so in the foreseeable future as there are considerable proven reserves of suitable coal.

New South Wales has the largest population of any State (4.500,000) and approximately 90% of its energy is generated in thermal stations, the annual coal consumption being 9 million tonnes (1975). Relatively high population densities and the degree of industrialisation combined to make New South Wales and Victoria the two States where effective Clean Air legislation was first introduced about 15 years ago.

Although Australian coals are usually low in sulphur (often less than 1%) only the black coals of New South Wales and Queensland are known to produce fly ashes which can be difficult to collect in cold precipitators. The fly ash from Victorian lignites and the sub bituminous South Australian coals can be effectively handled by precipitators having specific collecting areas of 40 to 60 $m^2/m^3/s$ whereas the black coal plants require up to 100% or more additional plate area.

This paper reviews the application of flue gas conditioning as an aid to precipitator performance in New South Wales with a passing reference to Queensland where, to date, it has been used operationally at one plant.

2.0 Coal Quality

The locations of coal sources and principal power stations are shown in figure 1. At this time, generation in the Sydney area is negligible, the comparatively small quantity of coal used there coming from within the northern or western coalfields whereas 20 years ago the Metropolitan stations imported coal from a large number of mines located in the several coal bearing areas. Table 1 summarises the characteristics of coal supplied to the major stations.

3.0 Fly Ash Characteristics

Table 2 gives typical chemical analyses of some Australian fly ashes, each ash being related to its parent coal by name. Considerable variability of constituents is found and therefore the stated values should be regarded as typical rather than absolute. Unless otherwise stated, any resistivities quoted are determined by Kevatron equipment, at, or about, the normal precipitator operating temperature and at a typical value of moisture in flue gas. All particle size determinations are by Bahco equipment.

The nature of most New South Wales coals is such that the fineness of fly ash more or less reflects the fineness of coal grind in the boiler milling equipment

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* Supervising Engineer, Power Design Group, Electricity Commission of N.S.W.

Table 1

Characteristics of Power Station Black Coal

Coal seam - Area - Stations -	Great Northern Central Coast Vales Pt. Munmorah	Tongarra/ Wangawilli Southern Tallawarra	Bayswater Hunter Valley Liddell	Lithgow Western Wallerawang	Callide Callide (Qld) Gladstone
Air dried basis					
Moisture	3.1	1.1	3.4	3.0	9.0 (1)
Ash	15.6	32.8	30.0	18.0	15.8
Volatiles	30.6	22.8	23.2	31.2	26.7
Fixed carbon	50.7	43.3	43.1	47.0	57.7
Heat Value MJ/Kg	27.80	22.6	21.24	25.84	25.74
Sulphur total	0.36	0.42	0.34	0.58	0.2
pyritic	0.00	-	0.11	-	0.12
sulphate	NIL	-	NIL	-	NIL
organic (diff.)	0.36	-	0.23	-	0.08
fixed	0.06	-	0.16	-	0.13
Carbon	67.7	55.9	53.7	65.6	67.2
Hydrogen	4.3	3.5	3.3	4.3	3.3
Nitrogen	1.3	1.2	1.2	1.5	1.0
Sulphur (organic)	0.4	0.4	0.2	0.7	0.1
Oxygen	7.6	5.2	8.2	6.9	12.6
Hargrove Index	47	65	51½	54	83½
Ash fusion temperatures °C					
Reducing atmosphere					
Deformation	1270	> 1400	1350		1390
Hemisphere	>1400	> 1400	> 1400		1550
Flow	>1400	> 1400	> 1400	>1600	>1600
Oxidising atmosphere					
Deformation	1320	> 1400	> 1400		1520
Hemisphere	>1400	> 1400	> 1400		1600
Flow	>1400	>1400	> 1400	>1600	>1600

(1) As Fired.

Table 2

Analyses of Typical Black Coal Fly Ashes

Coal Seam (Station)	Great Northern (Vales Pt.)	Tongarra/ Wangawilli (Tallawarra)	Bayswater (Liddell)	Lithgow (Wallerawang)	Callide (Q.A.L.)(1)
SiO_2	55.6	65.0	57.8	63.1	41.4
Al_2O_3	29.6	25.0	28.9	26.8	38.8
Fe_2O_3	4.53	7.24	4.71	0.59	10.15
TiO_2	1.41	0.60	1.10	0.88	1.94
Mn_3O_4	0.07	0.14	0.04	Nil	0.20
CaO	2.27	0.33	1.76	0.51	2.80
MgO	1.35	0.48	1.70	0.30	1.70
Na_2O	0.65	0.07	0.50	0.12	0.14
K_2O	2.98	1.01	1.27	2.52	0.08
P_2O_5	0.07	0.05	0.40	0.18	1.04
SO_3	1.00	0.20	1.35	0.30	0.16

(1) Queensland Alumina Ltd., Gladstone, Queensland.

4.0 Precipitator Installations

During the Second World War and the years immediately after, shortages of coal and of generating plant diverted attention from deficiencies of dust collecting plants and it took some time to accept the fact the early precipitators were quite unable to meet guaranteed performance. From about 1955 on, there has been a consistent effort to improve the collecting efficiency of existing equipment and to ensure that new plant performs adequately.

Table 3 lists N.S.W. precipitator installations ordered prior to 1958. The lack of understanding of the problems of fly ash collection from low sulphur coals, general at those times, is regrettably obvious when the specific collecting areas provided are related to the guaranteed efficiencies. As a consequence emission levels were up to 40 times the expected level.

Table 3

Precipitators Ordered Before 1958

Year in Service	Station/ Plant No.	Total Equivalent MW	Typical Operating Temperature °C	Nominal S.C.A. $m^2/m^3/s$	Guaranteed Efficiency %
1939/41	Bunnerong Nos. 19-22	140	180	8	N.A.
1951/56	White Bay Nos. 1-4	100	180	N.A.	98.08(1)(4)
1952/53	Pyrmont Nos. 1-2	100	180	40	96.0
1953	Bunnerong Nos. 23-24	60	150	40	97.0
1953/55	Pyrmont Nos. 3-4	100	180	40	97.5
1957/59	Wallerawang Nos. 1-4	120	360	16	99.0(1)(2)(3)
1958/60	Wangi Nos. 4-6	180	150	52	98.3(1)(3)(4)
1960/61	Tallawarra Nos. 5-6	200	140	54	99.37(1)(3)(4)
1961	Wallerawang Nos. 5-6	120	150	40	98.5(1)(4)
TOTAL = 25 plants (1100 MW)					

Notes (1) Includes precollectors
(2) Hot precipitators
(3) Now retired
(4) Gas conditioning available

Precipitators purchased for N.S.W. power stations since 1958 have, directly or indirectly, been sized from extensive pilot plant tests with the result that the performance of the new installations is significantly better than previously. (See Table 4).

Table 4

Precipitators Ordered after 1958

Year in Service	Station (Unit)	Equivalent MW	Nominal Temp °C	Design S.C.A. $m^2/m^3/s$	Guaranteed Efficiency %
1963	Vales Point (No. 1)	200	132	53	99.0 (4)
1964/65	Vales Point (Nos.2,3)	400	132	64	99.0 (4)
1966	Vales Point (No. 4)	275	132	74	99.0 (4)
1967/68	Munmorah (Nos.1,2)	700	138	96	99.0
1969	Munmorah (Nos.3,4)	700	138	78	99.0 (4)
1970/73	Liddell (Nos.1-4)	2000	Pri: 180 Sec: 115	73	98.5 (4)
1976	Wallerawang (No. 7)	500	115	112	98.5 (1)
1976	Wallerawang (Nos.5X, 6X)	240	150	86	99.0 (2)(3)(4)
1977/78	Vales Point (Nos.5,6)	1320	104	87	99.5
1981	Wallerawang (No. 8)	500	115	112	98.5 (1)

TOTAL = 18 plants (6835 MW)

Notes: (1) At maximum gas temperature
(2) With gas conditioning at M.C.R.
(3) Replacement units
(4) Gas conditioning available

With the exception of one plant where there was a small shortfall, all the operating plants listed in Table 4 have met or exceeded guarantee, some by substantial margins.

The replacement precipitators at Wallerawang (5X, 6X) are of special interest as they are pressurised units, installed between the I.D. fans and the stack. The existing collectors (mechanical on boilers 1 to 4, precipitators on boilers 5 and 6) will remain in service to protect the I.D. fans. It will be noted from figure 2 that the nominal gas temperature of 150°C is particularly adverse at Wallerawang. (Curve G). Unfortunately, intermittent operation of the station precludes use of water sprays for temperature modification.

5.0 Flue Gas Conditioning - Historical Background

The beneficial effects of flue gas conditioning were first demonstrated in Australia at Wangi and Pyrmont in the late 1950's. Whitehead (1) has reported experience with steam and/or water at Pyrmont in 1956/57. For reasons not fully understood at the time, the results were both variable and inconclusive. In addition, there was continual trouble with sprays blocking, hoppers plugging and fallout in ducts. These considerations led to water sprays being abandoned as the primary means of improving precipitator performance although some steam (or water) conditioning has had a secondary role up to the present time.

In 1960/61 SO_3 from a sulphur burning plant at Tallawarra dramatically improved the collection efficiency of the precipitators. Unfortunately, operating problems with the sulphur burner led to the trials being terminated after a short time. At about the same time, several precipitator manufacturers were operating large pilot plants at Wallerawang, Wangi and Pyrmont and some testing time was given to conditioning trials.

Because stabilised SO_3 is not commercially available in Australia, an experimental sulphuric acid plant was built and operated at Pyrmont in 1964. Modifications of this equipment were later used in other stations with varied success. In that year pilot plant studies with ammonia and with temperature modification were made at Tallawarra where exceptionally high resistivities ($> 10^{12}$ ohm metres) make cold precipitation most difficult. Operational conditioning with ammonia commenced at Tallawarra in 1965/66.

From 1965 on, gas conditioning has been increasingly used to boost the performance of undersized plants or plants where changed circumstances have resulted in excessive emissions. By February, 1976 more than 4,000 MW of boiler plant has been provided with gas conditioning facilities to aid precipitator performance when, and as, required.

6.0 Gas Conditioning Options

Flue gas conditioning can provide a means of raising the performance of a sub-standard precipitator to a satisfactory level or reducing the size of a new precipitator installation. A number of conditioning agents are distinctly unpleasant substances and all involve problems of installation, control, maintenance and add to operating charges.

The techniques tried in Australian power stations are listed below:

6.1 Gas Temperature Modification

While the broad relationship between temperature and resistivity is a matter of topical interest and has been referred to in many recent papers, the effect of gas temperature on the precipitability of fly ashes is of more direct concern than resistivity to operators and designers. Figure 2 shows the temperature sensitivity of several fly ashes from which it is seen that not all of these conform to the 'vee' form or something approximating a mirror image of a resistivity-temperature curve. Further, in those cases where a definite vee occurs, minimum collection efficiency is found at perhaps 60°C above the point of maximum resistivity.

In cases where temperature sensitivity is favourable, performance can be improved by suitably modifying gas temperature without other conditioning. This was proposed for use at Tallawarra but later abandoned as a primary corrective measure in favour of ammonia.

6.2 Water or Steam Conditioning

Extensive trials have been made with water and steam at several stations. Although stack emission was reduced in most cases the improvement was not sufficient and trials were discontinued after several months. The technique is worthwhile where a precipitator is marginally undersized or where gas temperatures are initially considerably above 100°C.

6.3 Phosphorous Pentoxide

Although P_2O_5 has been found to improve collection efficiency in pilot precipitators, its cost and nature has discouraged its use in full sized plants.

6.4 SO_3 Conditioning

Although sulphur trioxide is not commercially available in Australia, small quantities have been imported for pilot plant tests. Performance improved as injection rates were increased to about 60 p.p.m. $^V/V$, however with most fly ash a plateau is reached at about 30 p.p.m. and no improvement occurs with higher concentrations.

6.5 Sulphuric Acid

Figure 3 shows diagrammatically the sulphuric acid vapourising system employed at Pyrmont and other stations. Concentrated sulphuric acid passes into tubular Hastalloy grids carried above a bank of electric heating elements in a furnace of firebrick, the resulting vapour being injected into the boiler after the superheater. The system is comparatively simple and has a high availability with close attention. However, concentrated acid is not an attractive commodity in power stations and other techniques are preferred wherever possible.

In the developmental stage of acid vapourisers there was a problem with an occasional discharge of acidic agglomerates from station stacks which was countered by providing for neutralising by ammonia injection after the precipitator.

6.6 Ammonia

This is the most commonly used conditioning reagent in Australian applications although it is by no means universally (or equally) effective. Furthermore it has not been found possible to predict its effect with any certainty.

Figure 4 shows (in a simplified form) the anhydrous ammonia injection system at Liddell. Road tankers transfer liquid ammonia to the bulk storage tank from which it is drawn in the liquid phase. Passing through a water warmed vapouriser (drawing water from the C.W. discharge canal), the gas passes from the receiving area to the station by a graded line and is delivered to each of the four precipitators by a bus main. Each precipitator is provided with remotely operated valves arranged to allow the plant operators to select manually either '100%', '50%' or 'OFF' by push button operation from the main plant control rooms. Pre-set needle valves control the injection rate which gives a nominal concentration of 40 ppm $^V/V$ at full load. Below approximately 60% load the 50% setting is used. No injection is needed at loads less than at 40%.

The Liddell boilers have a divided flue gas system (provided to burn high moisture coal in the later life of the stations). Up to the 30% by weight of flue gas at temperature from 125°C to 200°C are handled by two of five parallel precipitator paths. The balance, at temperatures between 90 and 125°C, is treated by three separate parallel paths. Injection of ammonia takes place at a point between the four air heater outlets and the point of duct division into the fire precipitator paths. Figure 5 shows the arrangement of sparge pipes in relation to the ducts and also the details of these within the duct. When ammonia is not being injected, the vent valves (see Figure 4) allow a small flow of atmospheric air to be sucked into the ducts to inhibit jet plugging.

An interesting alternative to anhydrous ammonia is a steelworks by product, ammoniacal liquor (C.A.L.) which was readily available at low cost and was used at two stations. Almost the only constant thing about C.A.L. was the ammonia content, which ran at approximately 1%, and its general unpleasantness. With proper attention to storage and filtering it was found to be quite satisfactory and was extensively used at both stations.

Figure 6 shows the basic concepts of the C.A.L. system used on the 100 MW units of Tallawarra 'B' where its effectiveness was judged to be even better than anhydrous ammonia.

It is to be noted that all stations using ammonia inject it after the air heater and therefore the problem of air heater plugging does not arise unless ammonia leaks occur after boiler shut down. The only cases reported of build up on precipitator internals, or of difficult dust extraction from hoppers, are believed to have occurred when injection rates have been increased in attempts to offset the effect of zones or paths being out of service or when load reductions have not been followed by reduced injection rates.

6.7 Triethylamine

The discovery (or rediscovery) of triethylamine has been reported by Paulson (2) and is perhaps the most significant development in the field for many years. An expensive product of the petrochemical industry, the comparatively low concentration required can make it competitive with H_2SO_4 when the cost of post conditioning with ammonia is included. However, in the Australian price structure, it is rather more expensive than ammonia where that reagent is adequately effective.

Triethylamine has been used regularly at Wallerawang since October, 1975 after being tested there in a large pilot plant earlier in the year. It has also been used experimentally at a cement plant in New South Wales, at two stations in Western Australia with very satisfactory results.

At the time of writing (February, 1976) certain environmental aspects are still being studied and the extent of its future use depends on the outcome of these and on tests being carried out on other ashes.

7.0 Effectiveness of Conditioning

Reference to figure 2 shows the relative effect of temperature modification on a number of fly ashes. It should be noted that the curves demonstrate the effect of variations of flue gas temperature on precipitability (expressed as migration velocity) and are not some reciprocal function of resistivity. It is readily seen that prior knowledge of temperature response is helpful in determining whether an increase or decrease of temperature is likely to be helpful. (See Appendix for an explanation of Wk).

As earlier mentioned, steam or water injection may modify temperature as well as increasing the moisture content of flue gas. If both of these act in the same direction i.e. if the lowered temperature results in operation at a more favourable part of the temperature curve as well as reducing gas volume AND the added moisture further reduces resistivity, maximum benefit results. While there is a general tendency for added moisture (in flue gas) to aid precipitation, this response does not appear in some cases, probably because of the particular chemistry of the surface layer of dust particles (3).

The effectiveness of ammonia, sulphuric acid and triethylamine as conditioning agents are shown in fig. 7 from which it is at once apparent that sulphuric acid (or SO_3) is generally more effective than ammonia and that triethylamine appears to be superior to both. However, it will also be noted that the responses of fly ashes to all three reagents are quite variable. While there is little doubt that SO_3 (or H_2SO_4) reduces resistivity and that therefore it might be argued that a competently designed and well maintained precipitator, sized to achieve high collection efficiencies with high sulphur coal should be able to achieve the same efficiency if highly resistive fly ash if conditioned with SO_3, Australian data suggest that other factors may be involved.

Some evidence of the possible significance of particle size has recently emerged at Wallerawang where pilot precipitator tests during 1975 gave an improvement of 30% in W with dust of median size 13 - 15μm from the B station, whereas A station dust, following mechanical collectors and having a median size of 6-8μm, showed little response. While one explanation is that the mechanical collectors removed the chemically reactive portion of the dust, a review of earlier data showed a similar lack of sensitivity in dusts below about 10μm (median). Further study is indicated.

Notwithstanding the marked improvement in electrical characteristics with Tallawarra fly ash, as earlier reported (4), there was no measurable change in resistivity. This finding is consistent with that of Dismukes (5) at several T.V.A. stations. However, although the Tallawarra ash showed a quite spectacular response to ammonia, this was less rapid than reported by Dismukes at Gallatin.

8.0 Conditioning Rates

The optimum injection rate for any reagent is obviously the minimum quantity capable of producing the desired improvement. Unfortunately, there is considerable variation in the optimum rate of injection of most conditioning agents and, further, as mentioned earlier, it is possible to over condition.

With SO_3 and H_2SO_4, the preferred rate has quite often been 20 - 30 ppm V/V. Figure 8 shows the response to sulphuric acid vapour at Wangi.

The optimum rate for anhydrous ammonia and CAL (Section 6.6) is very variable; at one time Tallawarra was conditioning at the rate of 200 to 300 ppm whereas the Gladstone plant of Queensland Alumina operates at only 22 ppm (6) Fig.8 also shows the effect of varying the rate of anhydrous ammonia injection at Wangi.

Reference was earlier made to the potency of triethylamine. Although insufficient work has been done to determine whether this also varies in its effect from ash to ash, it appears that the optimum rate is less than 10 ppm by weight.

At the risk of stating the obvious, the need to achieve effective mixing of any conditioning agent with flue gas sufficiently far upstream of the precipitator to provide adequate residence time cannot be too strongly emphasised. In retrospect, it is probable that some of the plants requiring high injection rates were suffering from the effects of poor distribution, a situation which arises if nozzles are blocked or if the basic design is poor.

Where a carrier gas is used (e.g. hot air) to introduce a conditioning agent into flue gas ducts, mixing will be assisted if close attention is given to achieving densimetric similarity between the carrier gas plus the gaseous reagent and flue gas. If a reasonable match is achieved at M.C.R. and satisfactory results obtained, performance at lower loads is also likely to be good without modifying the carrier gas temperature.

9.0 Economic Considerations

The capital costs of conditioning plant reflect the basic differences between the storage and injection systems appropriate to each reagent. The properties of $H_2 SO_4$ permit use of carbon steel in some parts of the installation but also require the use of more exotic materials to handle spills and leaks when the acid may be diluted. Those considerations and the need for post neutralising ammonia injection make the capital cost of sulphuric acid plants the highest of those so far used in Australia, an indicative cost being $A 0.12/KW of boiler plant. The comparative simplicity of anhydrous ammonia systems reduces the prime cost to about $A 0.05/KW and it is probable that triethylamine installations may be cheaper still.

The approximate cost of gas conditioning, based on February 1976 prices is given in Table 5.

Table 5

Conditioning Costs
(Feb. 1976 prices)

Conditioning Agent	Price $A tonne	Cost cents/Mwhr	Injection Rate
NH_3	240	1.0	20 ppm V/V
$H_2 SO_4$	60	3.0	20 ppm V/V plus neutralising ammonia
$N (C_2 H_5)_3$	1200	5.0	10 ppm W/W

10.0 Precipitator Capital Costs

Reference was earlier made to the fact that in New South Wales gas conditioning has been provided on more than 4000 MW of precipitators. Although at first sight, this would appear to be an indictment of cold precipitators, this is not so, as 68% of conditioning installations are on the older plants listed in Table 3 and of the plants listed in Table 4, four installations are conditioned because emission limits were tightened between ordering and commissioning and one because it has a percentage of plate discharge electrodes which have not met performance expectations.

Some comments on the estimated relative Australian costs of hot and cold precipitators (conditioned and unconditioned) may be of interest. 1976 prices for a large (87,000 m^2) cold precipitator of 20 fields are approximately $A 12/kW, including ductwork from air heater to stack. Assuming that a hot precipitator costs of the order of 20% more than a cold unit of the same plate area and with an allowance of 70% for additional gas volume, a hot precipitator becomes competitive when the total migration velocity (W_k) approaches twice the cold value. This corresponds to a $\sqrt{2}$ or 40% increase in e.m.v. (w). On the Australian scene, only two fly ashes are known to offer the possibility of lower cost hot precipitators and although the specifications admitted hot plants, they were not offered.

With the exception of those few cases where fly ash is impossible to collect in any reasonably sized cold precipitator, the three precipitator options are cold, cold conditioned and hot. The relative plate areas and relative cost of each of these are summarised in Table 6.

Table 6

Comparison of Precipitator Alternatives For 99.5% Collection Efficiency

	Wk Ratio	Specific Collecting Area $m^2/m^3/s$		Ratio of Plate Areas		Ratio of Capital Costs (4)	
		Min	Max	Min	Max	Min	Max
Hot (1)	2	40	80	1.0	2.0	1.0	1.8
Cold (2)	1	80	160	1.2	2.4	1.0	1.8
Cold (2)(3) (Conditioned)	2	40	80	0.6	1.2	0.5	0.9

Note:
(1) Performance data from U.S. plants.
(2) Performance data from Australian plants.
(3) Assume Wk increased by not less than 100% with conditioning.
(4) Not including cost of conditioning plant.

It is seen from Table 6 that conditioning offers the prospect of reduced capital costs in cases where there is a high level of confidence in the response of fly ash to conditioning and that use of hot precipitators does not necessarily confer a cost advantage. It is readily conceded that other considerations such as availability and O and M charges can have considerable impact in the long term.

11.0 Conclusions

Gas conditioning in one form or another can be considered a reliable means of improving precipitator performance in some circumstances. Prediction of optimum concentration or the extent of improvement is seldom, if ever, possible without some form of prior testing. Care in positioning and servicing the injection nozzles is necessary as is attention to controlling injection rate, although load modulation should not be required.

Installation of undersized precipitators with gas conditioning in conjunction with new boiler plant is not recommended without adequate supporting evidence of the likely effectiveness of conditioning with all possible coals. However, the order of saving in capital expenditure possible, where an adequate confidence level can be established, suggests that investment in appropriate prior research and/or pilot plant testing may be worthwhile.

Triethylamine offers considerable promise as a highly effective reagent although its chemical similarity to ammonia may be shown to make it unsuitable for some fly ashes.

Gas conditioning is comparatively inexpensive in cases where it produces a reasonable improvement at low concentrations; even where concentrations of 50 to 100 ppm are required to effect the desired improvement in collection efficiency, conditioning may well be cheaper than alternative actions.

Although resistivity is an important factor in determining precipitator performance, it is not the only consideration and therefore recourse to use of appropriately sized pilot plants is recommended whever possible.

Size for size, hot precipitators are more expensive than cold and, further, attract additional cost penalties unless the boiler layout is suitably modified at the design stage. On the evidence available, it is doubtful if all claims as to their merits in low sulphur coals can be sustained. In some instances, a significantly cheaper conventional unit will perform as well or better. There are also cases where hot plants will not only be cheaper than unconditioned cold units but where, in fact, they provide the only technically acceptable alternative.

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APPENDIX

Total Average Migration Velocity - Wk

Extensive use is made of "total average migration velocity" (omega K or Wk) as distinct from "effective migration velocity "(e.m.v.) as used in the classical Deutsch formula.

$$E = 1-e^{-\frac{A}{Q}W} \quad - (1)$$

where E = collection efficiency
A = total active surface of collecting electrodes
Q = gas quantity in unit time
e = 2.7182

(Note that $\frac{A}{Q}$ = specific collecting area or S.C.A.)

It is frequently found in commercial precipitators that the W in the Deutsch formula decreases as the gas volume decreases and increases as the gas volume increases. Because W is not independent of $\frac{A}{Q}$ its use to transpose precipitator performance will give grossly misleading results. For example, if the specific collecting area of a precipitator operating at 98% is increased by 20% the emission, if computed by the Deutsch W, will be halved; in reality a 25% reduction is about all that can be expected. Conversely, use of the Deutsch formula to compute the effect of a 10% increase in gas volume indicates a 40% increase in emission whereas a properly designed and correctly operating precipitator should limit the increase to about 15%.

Several variants of the Deutsch formula have been put forward over the years. The one almost universally used in the Australian power industry is that developed by the Swedish company of Svenska Flaktfabriken and published by Matts and Ohnfeldt *.

$$E = 1-e^{-(\frac{A}{Q}\,Wk)^{k}} \quad - (2)$$

where Wk = total average effective migration velocity
K = $0.4 < K < 0.6$ (usually 0.5)

The total average effective migration velocity (Wk) has the advantage that it is independent of specific collecting area and therefore serves very well to predict the effect of adding (or removing) internal hardware or of varying gas velocity. It has been found that well designed and maintained precipitators invariably conform closely to a Wk relationship.

Wk serves as a performance index or figure of merit and is also useful whe when comparing flyashes collected in precipitators of similar design or when evaluating precipitators of different designs.

Comparison of equations (1) and (2) at once shows that

$$Wk = \frac{A}{Q}W^2 \quad \text{and} \quad W = \sqrt{\frac{Wk}{\frac{A}{Q}}}$$

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* Matts S., Ohnfeldt P.Q.: "Efficient Gas Cleaning with S.F., Electrostatic precipitators" S.F. Review, 1964.

from which it is seen that

$$\frac{Wk1}{Wk2} = \left(\frac{W1}{W2}\right)^2 \text{ and } \frac{W1}{W2} = \sqrt{\frac{Wk1}{Wk2}}$$

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Figure 9 shows a standard precipitator performance plotting sheet. This is one of several sizes ranging from 380 mm x 340 mm down and scaled in either metric or English units as desired. The vertical scale is efficiency plotted as ln(-ln(1-E)) and the horizontal scale is $\log_{10}$ (specific collecting area). Superimposed are two further scales, the more steeply inclined being the Deutsch 'W' and the less steeply inclined Wk. The format is different from that proposed by Matts and Ohnfeldt (figure 10) and is a development of the still commonly used Efficiency (log) vs SCA (linear) presentation (see figure 11) which has the disadvantage of representing precipitator performance by a curve.

By way of illustration, actual test points from an extensive series of pilot tests at Wallerawang in 1969 are plotted on each of figures 9, 10 and 11, together with the related performance bands. The versatility of the presentation shown in figure 9 is self evident, as is the validity of the Wk concept.

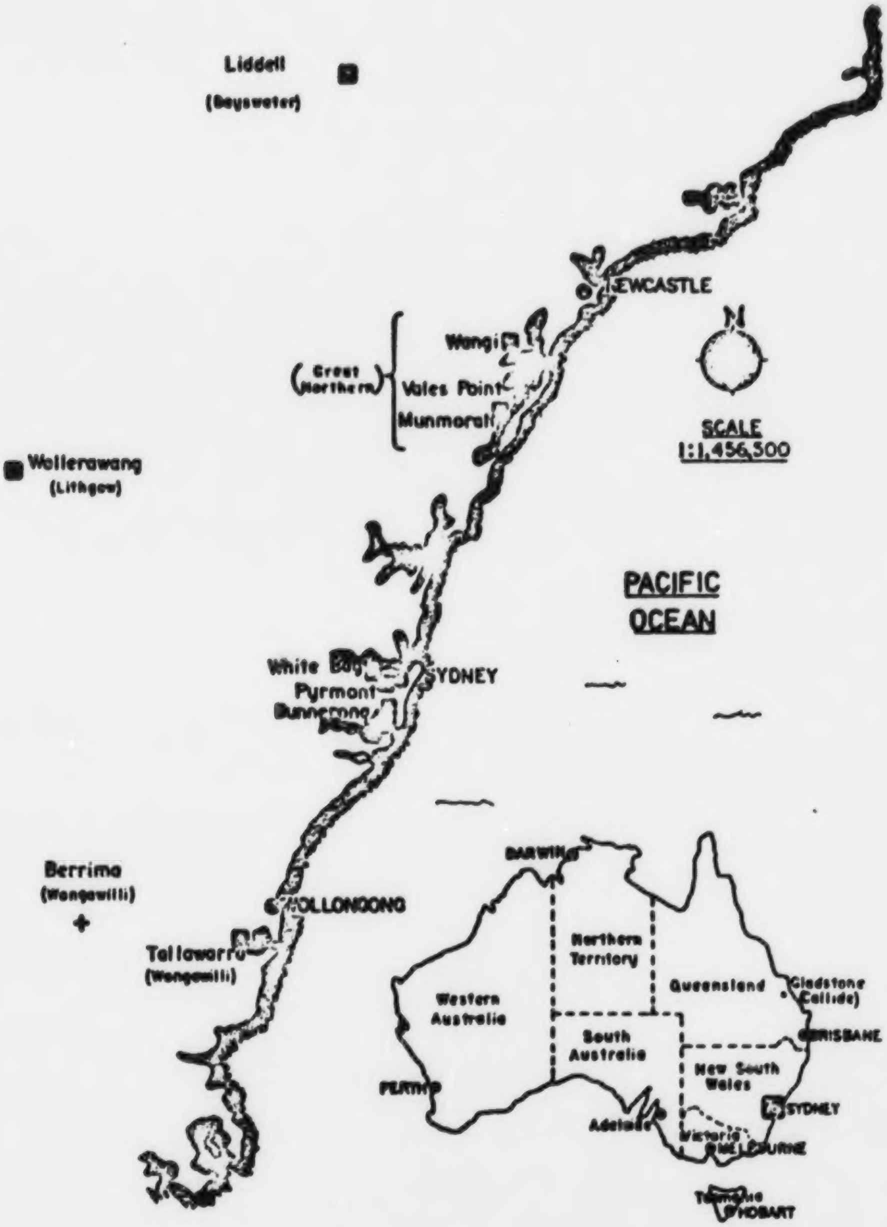

Figure 1. Location of Power Stations and Coal Fields (Coal fields in brackets).

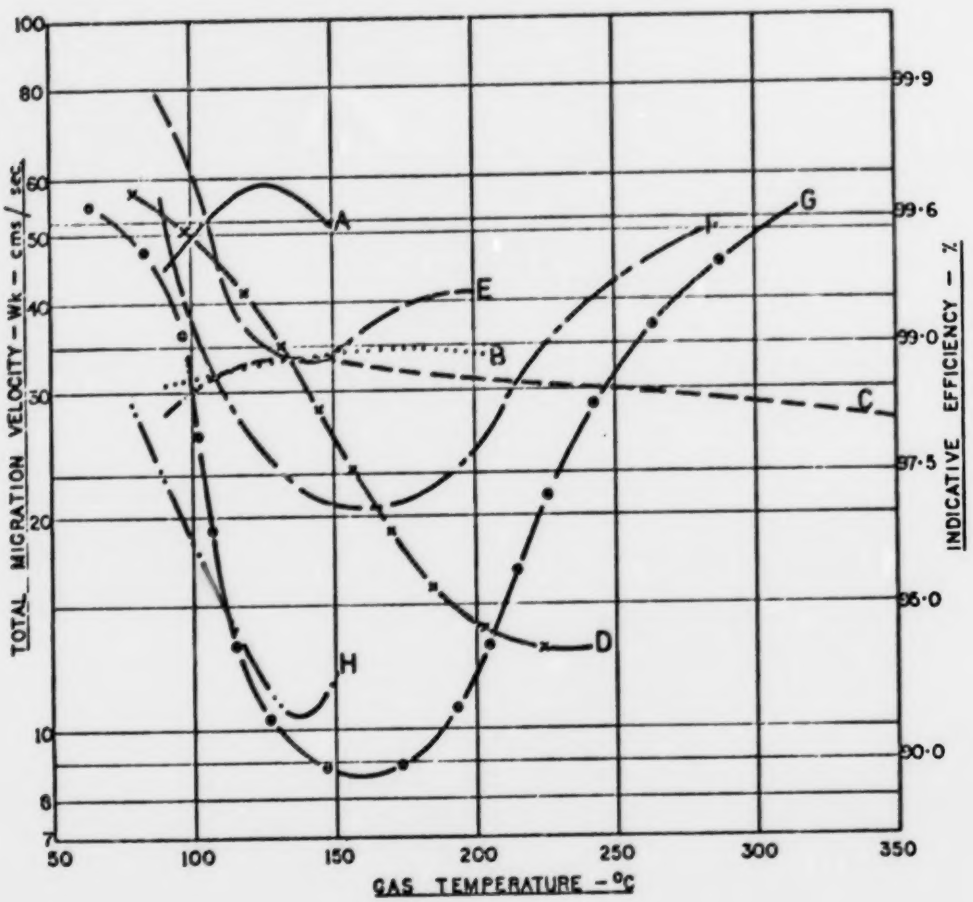

Figure 2. Temperature Sensitivity of Some Flyashes.

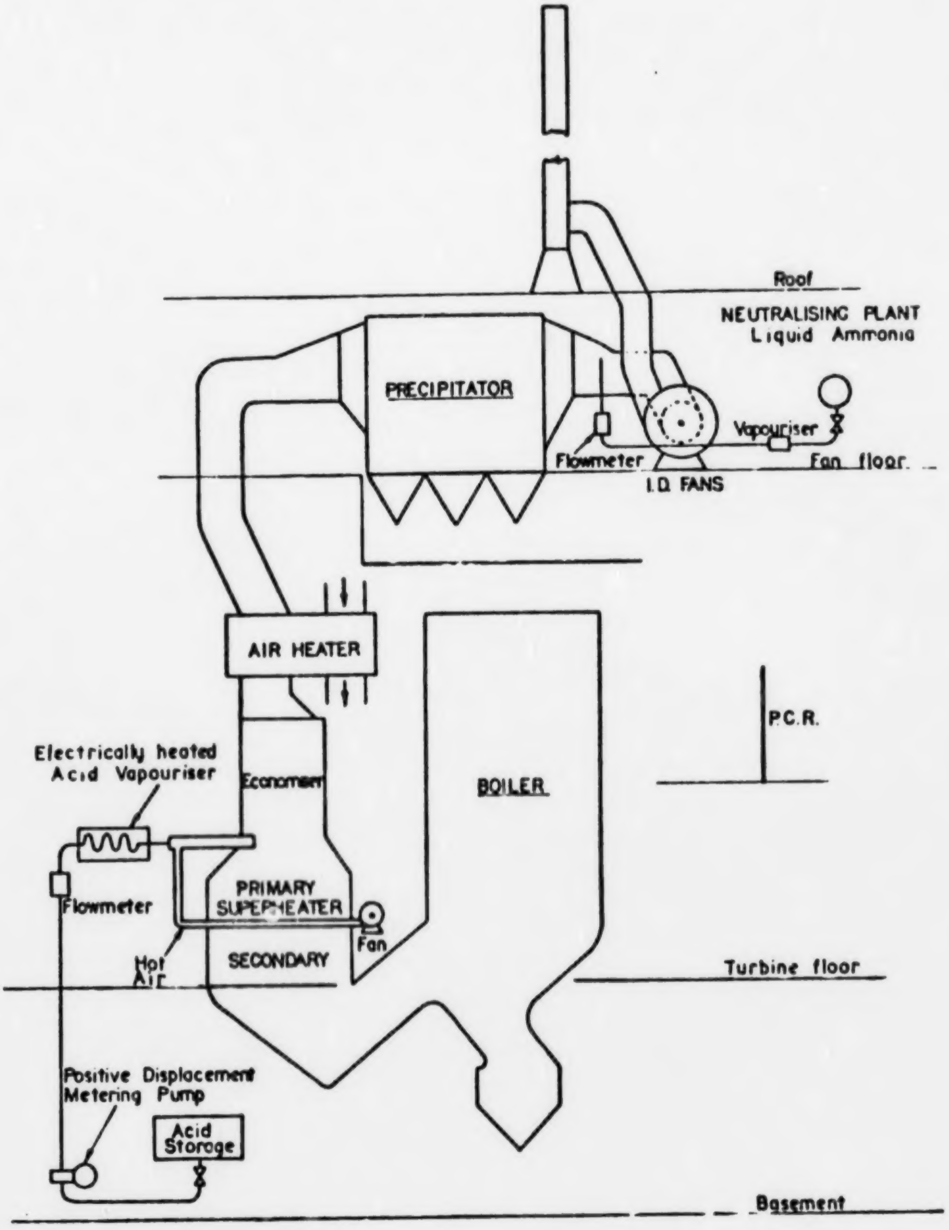

Figure 3. Sulphuric Acid Injection System Pyrmont.

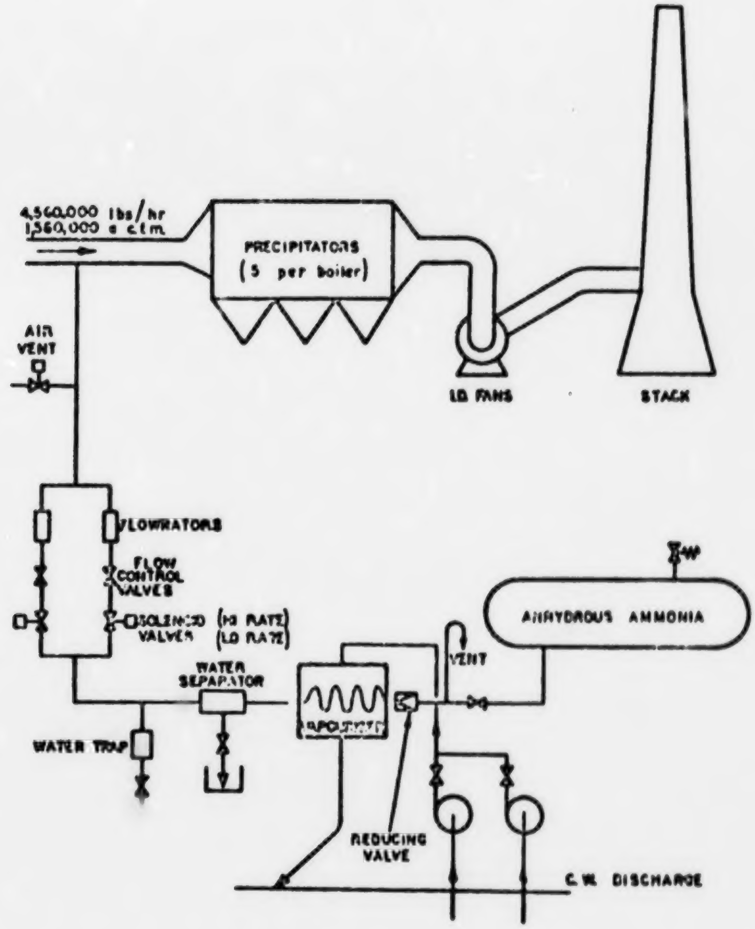

Figure 4. Liddell Ammonia Conditioning System
(diagrammatic only)

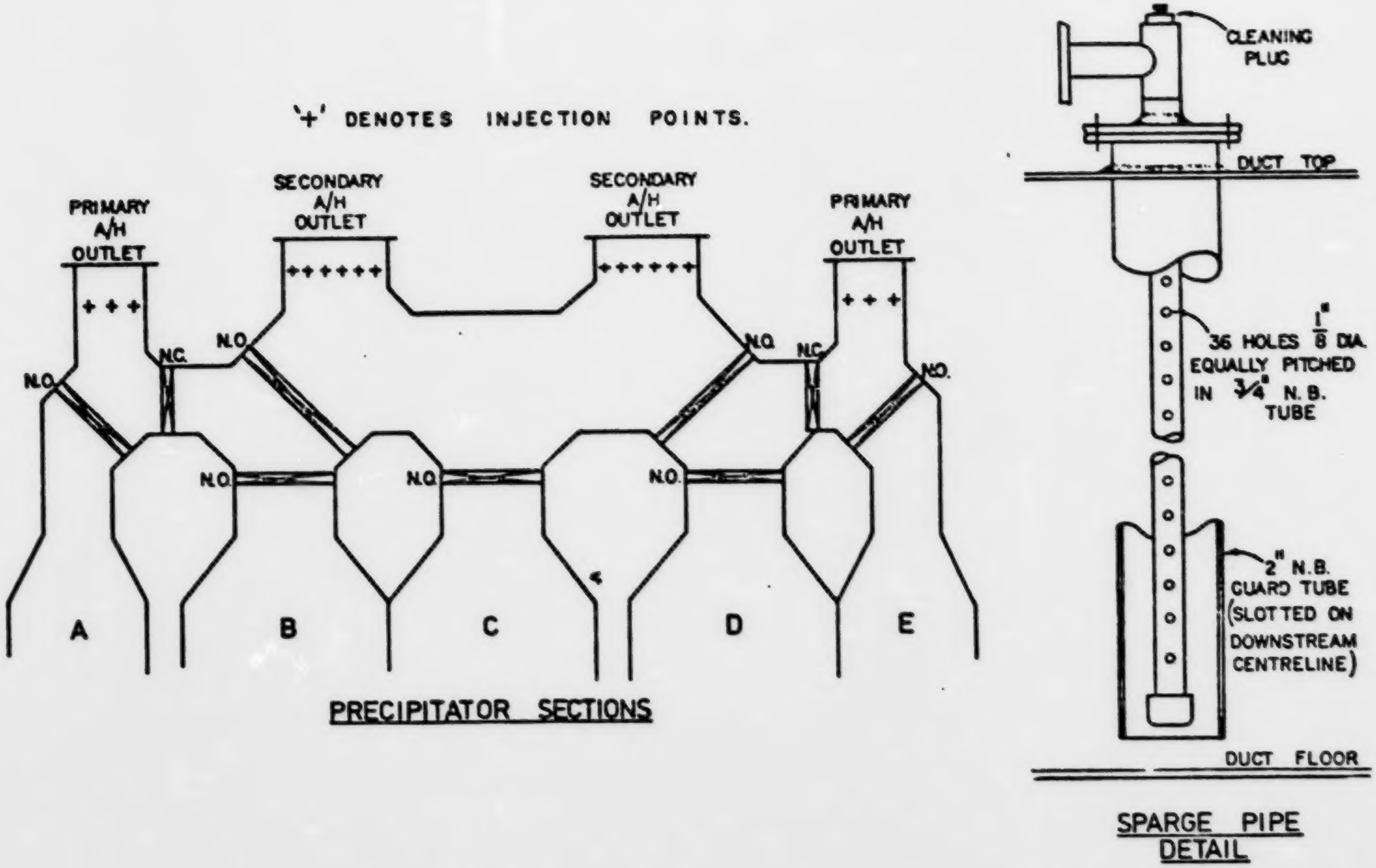

Figure 5. Liddell: Location of Sparge Pipes.

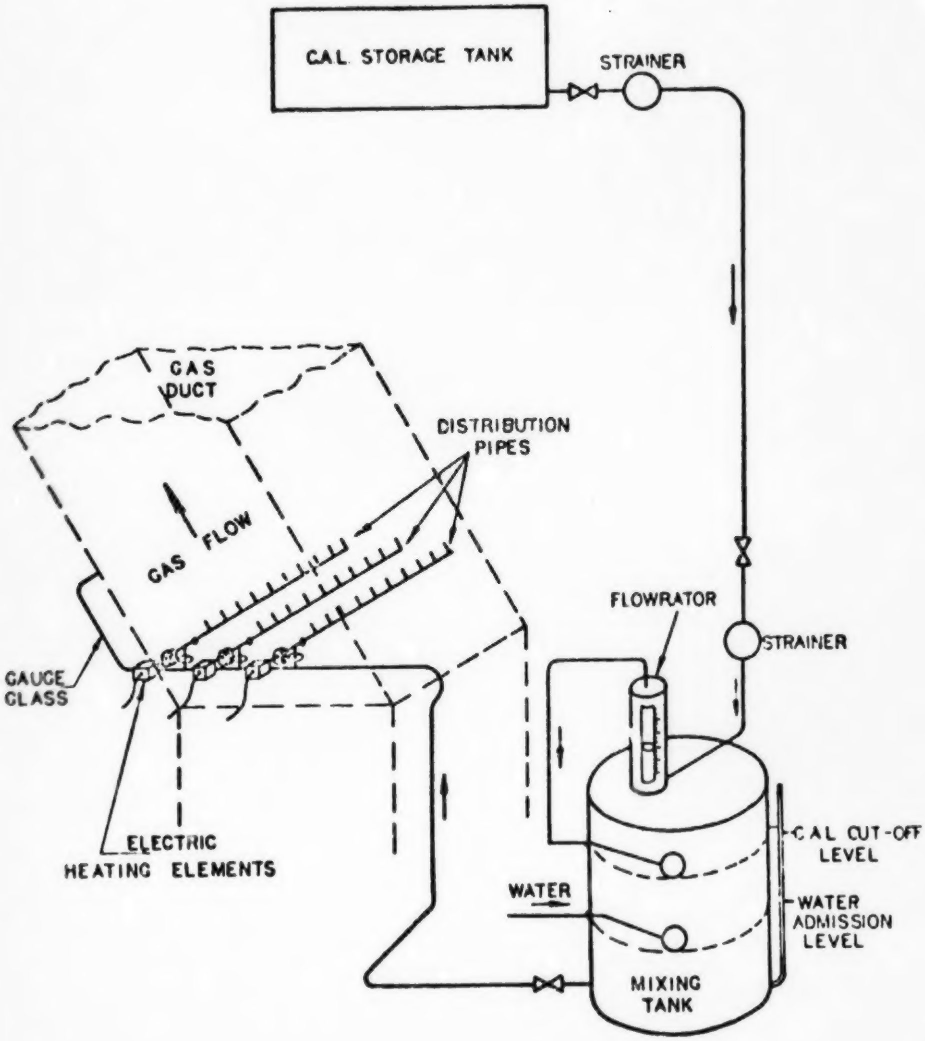

Figure 6. Tallawarra Power Station Cal. Gas Conditioning Plant (diagrammatic arrangement).

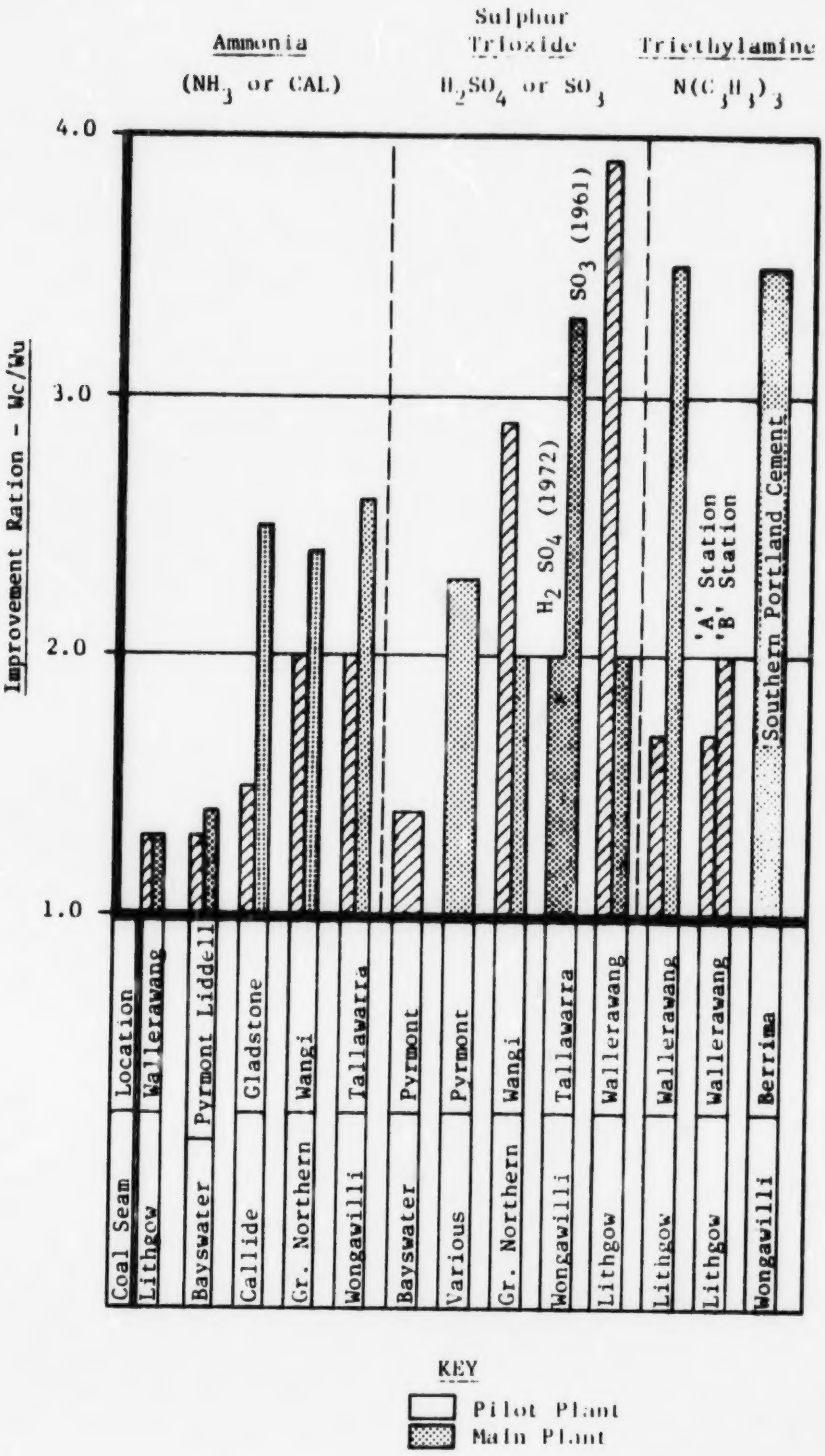

Figure 7. Selected Improvement Ratios from Conditioning.

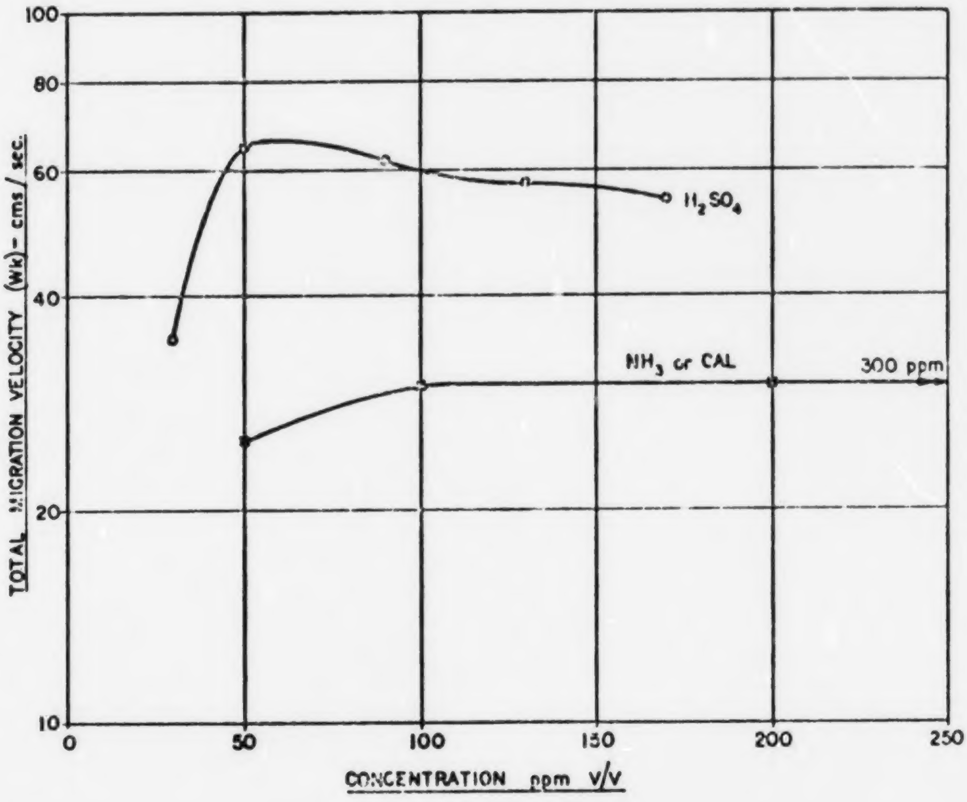

Figure 8. Wangi – 1972 Optimum Conditioning Rates.

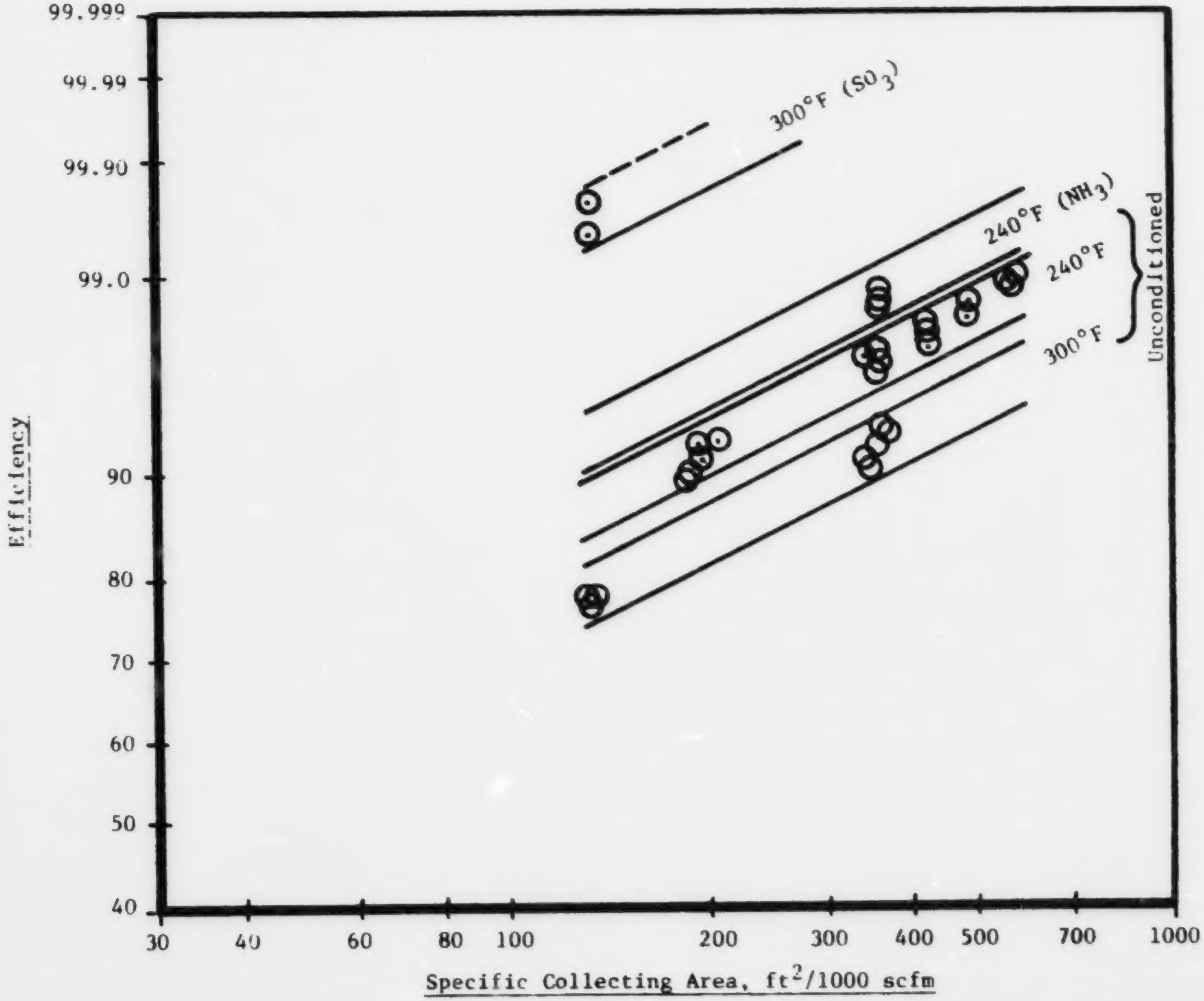

Figure 9. Effect of Conditioning Agents

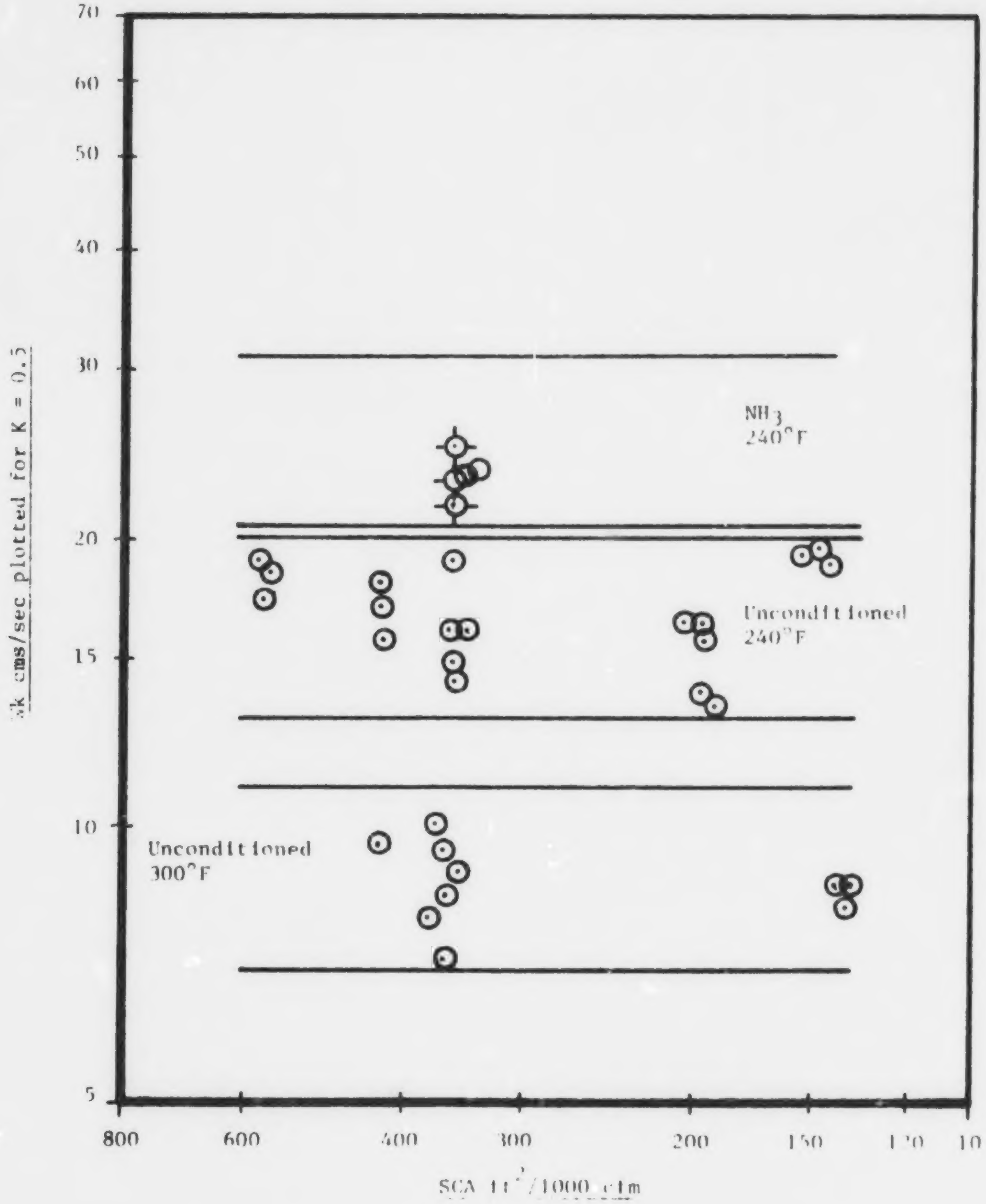

Figure 10. Relation Between Specific Collecting Area and Migration Velocity

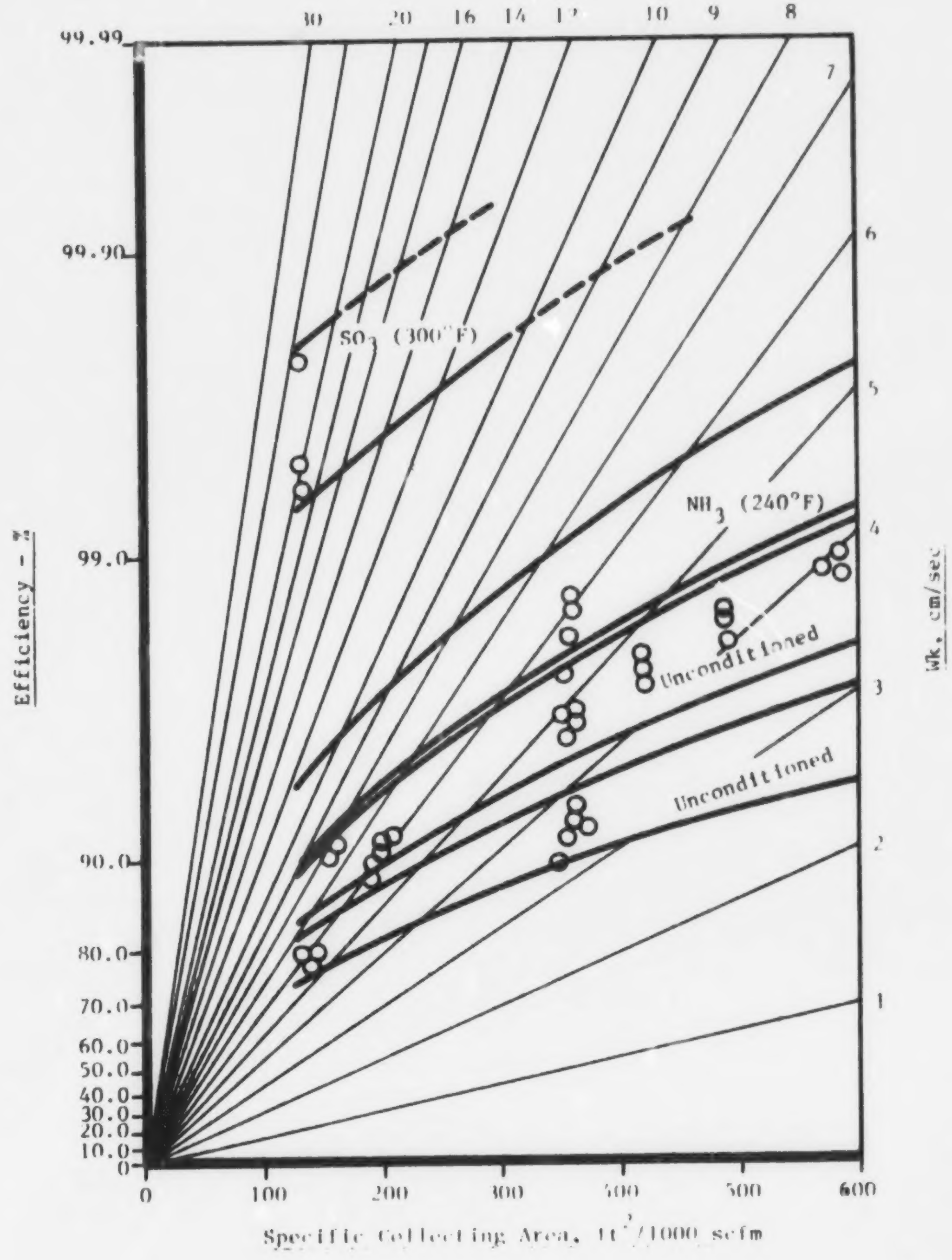

Figure 11. Efficiency Plot for Several Conditioning Agents

MATHEMATICAL MODELLING OF FINE PARTICLE COLLECTION BY ELECTROSTATIC PRECIPITATION

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ABSTRACT

A mathematical model is described which calculates collection efficiency in an electrostatic precipitator as a function of particle size and operating conditions. The model takes into account the dynamics of particle collection by determining the electric field, particle charge, and removal efficiency as functions of position along the length of the precipitator. The model contains procedures for estimating collection efficiency losses caused by non-uniform gas velocity distributions, gas bypassing the electrified regions, and particle reentrainment. Comparisons of the model predictions with experimental data obtained under essentially idealized conditions in a laboratory scale precipitator indicate that the model provides a reasonable basis for describing the precipitation process. Model predictions are compared with field data obtained from precipitators used to collect fly ash and the results are discussed.

MATHEMATICAL MODELLING OF FINE PARTICLE COLLECTION BY ELECTROSTATIC PRECIPITATION

INTRODUCTION

The techniques used in the sizing of electrostatic precipitators for various applications differ with the manufacturer and the application area. These techniques may be classified into two general categories:

1) The development of statistically-based relationships between precipitator performance and the properties of the dust or fuel, or the operating parameters of the precipitator. Usually this approach involves a correlation of a precipitation rate parameter which is used as an empirical factor in the Deutsch-Andersen equation.
2) The use of a pilot plant and prior experience for determining the collection area required for the design efficiency at a given application.

Although these techniques can be effective, they can also result in serious design errors if improperly used. Furthermore, in order to design a precipitator for a new installation, it is necessary to either have extensive data for similar applications, or to conduct a test program with a pilot unit. A preferable approach is to develop theoretical relationships to the point that precipitator design can be performed by calculation from theoretical principles and the properties of the dust to be collected. This approach offers the potential for increased confidence in design and in cost savings by preventing undersizing on the one hand and over-sizing on the other. The reliability of predictions obtained from such an approach is subject to the extent to which certain fundamental parameters are known, the degree to which the theoretical relationships describe precipitator operation, and the accuracy with which the factors that correct for non-ideal conditions can be modelled and determined. At present, efficiency losses due to non-ideal conditions are accounted for only by estimation procedures in which assumed values of the descriptive parameters are normally used.

In order to model particulate collection by the electrostatic precipitation process from first principles, it is necessary to (1) calculate electric fields as a function of position and operating conditions, (2) calculate particle charge as a function of time, electrical conditions, and particle size, and (3) mathematically represent the process by which charged particulate are transported from the bulk gas stream to the collection electrode. The mathematical model described in this paper includes these operations and predicts collection efficiency as a function of particle size and operating conditions in electrostatic precipitators.

Comparisons of the model predictions with experimental fractional efficiency data obtained in a laboratory precipitator indicate that the model provides a reasonable basis for describing the precipitation process. Application of the model to simulate conditions existing in full-scale and pilot-scale precipitators used to collect fly ash leads to results which are in reasonable agreement with experimental field measurements. In certain cases, however, part of the ideally calculated fractional efficiency curve lies below the measured curve. This type of discrepancy can be attributed in part to certain unmodelled and approximately modelled processes which are important under certain conditions in precipitators.

DESCRIPTION OF THE MATHEMATICAL MODEL

Ideal Calculation of Particle Collection Efficiency

The details of a mathematical model for electrostatic precipitation and a computer program which performs the various operations are described elsewhere in the literature.[1,2,3] In the following, a brief description of the physical processes is given, and the mathematical framework of the model is developed.

The first fundamental step in the model is to determine the electric field existing in the precipitator. An accurate calculation of the electric field is important because of its influence on particle charging and particle collection. Although the electric field in a precipitator is a function of position, the value of the electric field used in the particle charging calculations is approximated by the average value between the discharge and collection electrodes. In order to calculate the velocity of charged particles near the collecting electrode, however, it is necessary to compute the local electric field values in this region of space.

In the model the electric field is determined as a function of position by the numerical relaxation technique suggested by Leutert and Böhlen.[4] Using this technique, Poisson's Equation and the current continuity equation are solved simultaneously and subject to the existing boundary conditions. These two equations can be written in two dimensions in discrete form as:

$$\frac{\Delta^2 V}{\Delta x^2} + \frac{\Delta^2 V}{\Delta y^2} = -\frac{\rho}{\varepsilon_0} \qquad (1)$$

and

$$\rho^2 = \varepsilon_0 \left(\frac{\Delta V}{\Delta x} \frac{\Delta \rho}{\Delta x} + \frac{\Delta V}{\Delta y} \frac{\Delta \rho}{\Delta y} \right) , \qquad (2)$$

where

V = potential, volts

ρ = space charge, coul/m^3

x = distance perpendicular to gas flow from wire to plate, m

y = distance parallel to gas flow from wire to wire, m.

The numerical solutions to Equations (1) and (2) are obtained by iterating on a grid of potential and space charge density until convergence is obtained. In the iteration procedure, the space charge density at the wire is adjusted until solutions are found that satisfy experimentally measured boundary conditions of applied voltage and average current density at the plate. The major approximation in the method is the assumption that the motion of all charge carriers can, on the average, be described by a single "effective mobility." In addition, the "effective mobility" is assumed to be independent of position in formulating Equation (2). In previous work,[3] it has been shown that this technique yields theoretical electric field distributions which agree well with existing experimental data in the literature.

The introduction of a significant number of fine dust particles or a heavy concentration of large particles into an electrostatic precipitator significantly influences the voltage-current characteristics of the interelectrode space. Qualitatively, the effect is seen by an increased voltage for a given current compared to the dust-free situation. The increased voltage results from the lowered "effective mobility" of the charge carriers which occurs as the highly mobile gas ions are bound to the relatively slow dust particles, thus creating a "space charge". It is desirable to determine the space charge resulting from dust particles because this quantity influences the electric field distribution, especially near the collecting plate where, for the same current, it is raised above the dust free situation. Also, the "space charge" is a function of position along the length of the precipitator.

It has been shown[3] that the effect of "space charge" on the electrical conditions can be estimated by using an "effective mobility" given by:

$$b_e = b_i \left(\frac{j_T}{200\ j_P + j_T} \right) , \tag{3}$$

where

b_e = effective mobility, $m^2/(\text{volt-sec})$

b_i = ionic mobility, $m^2/(\text{volt-sec})$

j_T = total current density at the plate, amp/m^2

j_P = particulate current density at the plate, amp/m^2.

The derivation of Equation (3) is based on the assumptions that (1) free electron current can be ignored, (2) the current carried by the particulate is insignificant, (3) particle mobility can be treated as independent of size and on the average, $b_i = 200\ b_P$, and (4) the particles instantaneously acquire a limiting charge given by the "saturation charge" from field charging theory. The quantity j_T is measured experimentally and j_P can be determined for a given incremental length of the precipitator from an estimated or calculated overall mass efficiency.

In a given incremental length of the precipitator, the effect of "space charge" on the electrical conditions can be determined by solving Equations (1) and (2) subject to the condition that:

$$j_T = b_e \rho_P E_P , \tag{4}$$

where $\rho_P E_P$ is the product of the space charge density and electric field at the plate averaged over the number of grid points used in the direction of gas flow.

The second fundamental step in the model involves the calculation of particle charge as a function of particle size, residence time, and precipitator operating conditions. Accurate values of particle charge are necessary because the electrical force which causes the particle to migrate towards the collection electrode is directly proportional to the charge on the particle. In the model particle charge is calculated using the charging rate equation developed by Smith and McDonald[5] given by:

$$\frac{dq}{dt} = \frac{N_0 e b q_s}{4\varepsilon_0}\left(1 - \frac{q}{q_s}\right)^2 + \frac{\pi a^2 e \tilde{v} N_0}{2}\int_{\theta_0}^{\pi/2} \exp - \left[\left(\frac{qe(r_0-a)}{4\pi\varepsilon_0 a r_0 kT}\right.\right.$$

$$\left.\left. + \frac{[3ar_0{}^2 - r_0{}^3\ (K+2) + a^3\ (K-1)]}{kTr_0{}^2(K+2)}\right)\right] \sin\theta d\theta$$

$$+ \frac{\pi a^2 e \tilde{v} N_0}{2} \exp\left(- \frac{qe}{4\pi\varepsilon_0 akT}\right) , \qquad (5)$$

where

$$q_s = 4\pi\varepsilon_0 E_0 a^2\left(1+2\ \frac{K-1}{K+2}\right) ,$$

$$\theta_0 = \arccos\left(\frac{q}{qs}\right) ,$$

and

q = instantaneous charge on the particle, coul

q_s = saturation charge, coul

θ = azimuthal angle in a spherical coordinate system with origin at the center of the particle, radians

θ_0 = maximum azimuthal angle for which electric field lines enter the particle, radians

N_0 = free ion density, number/m^3

e = electronic charge, coul

ε_0 = permittivity of free space, $coul^2/(N\text{-}m^2)$

E_0 = average electric field between the electrodes, volt/m

b = ion mobility, m^2/(volt-sec)

$\tilde{v}$ = mean thermal speed of ions, m/sec

a = particle radius, m

k = Boltzmann's constant, J/°K

T = absolute temperature, °K

t = time, sec

K = dielectric constant of the particle

r_0 = radial distance along θ at which the radial component of the total electric field is zero, m.

Equation (5) represents the sum of three charging rates corresponding to three different charging regions on the particle surface. The first term corresponds to a region of the particle surface where field charging occurs; the second, to a region where field-enhanced diffusion occurs; and the third, to a region where diffusion occurs with the effects of the applied electric field neglected as an approximation. The charging rate given by Equation (5) approaches the results obtained from the classical field charging equation for large particles and high electric fields and reduces to the classical diffusion charging equation in the absence of an applied electric field.

The agreement between the results predicted by Equation (5) and Hewitt's[6] experimental data for dioctyl phthalate droplets is within 25% over the entire range of data that is available and is within 15% for practical charging times in precipitators. The agreement between the theory and the experiment over the particle size range 0.1-1.3 μm and a wide range of electric field strengths indicates that Equation (5) should predict closely particle charge as a function of time in the precipitator model.

The third fundamental step is the calculation of particle collection efficiency. Once the particle charge and the electric field adjacent to the collecting electrode are calculated, the electrical drift velocity, or migration velocity, resulting from the coulomb and viscous drag forces acting upon a suspended particle can be obtained. For particle sizes in the size range of interest, the time required for the particle to achieve the steady-state value of velocity is negligible, and the migration velocity is given by:

$$w = \frac{qE_pC}{6\pi a\mu} , \qquad (6)$$

where

w = migration velocity of a particle of radius a, m/sec
E_p = electric field adjacent to the collecting electrode, volt/m
C = Cunningham correction factor
μ = gas viscosity, kg/(m-sec).

Gas flow velocities in most cases of partical interest are between 0.60 and 1.8 m/sec, while theoretical migration velocities for particles smaller than 6.0 μm are usually less than 0.3 m/sec. The path of these smaller particles therefore tends to be dominated by the turbulent motion of the gas stream in the interelectrode region. The classical equation for describing particle collection in electrostatic precipitators under turbulent flow conditions was derived by Deutsch[7] and gives collection efficiency as a function of gas volume flow, collection area, and migration velocity in the form:

$$\eta = 100 \, [1 - \exp(-A_p w/Q)] \quad , \qquad (7)$$

where

η = collection efficiency of a particle of radius a, %
A_p = collecting area, m^2
Q = gas volume flow, m^3/sec.

The assumptions on which the derivation of Equation (7) is based are discussed in detail elsewhere in the literature and will not be restated here.[1,8] Of greatest concern in the modelling of the precipitation process is the assumption that gas turbulence provides sufficient mixing to establish a uniform particle concentration at any cross section of the precipitator. The validity of this assumption is most important in determining theoretical collection efficiencies for fine particles.

The mathematical model uses the Deutsch Equation to predict the collection fraction, $\eta_{i,j}$, for the i-th particle size in the j-th incremental length of the precipitator. Thus, the Deutsch Equation is applied in the form:

$$\eta_{i,j} = 1 - e^{-W_{i,j} A_j/Q} \quad , \qquad (8)$$

where $W_{i,j}$ (m/sec) is the migration velocity of the i-th particle size in the j-th increment, and A_j (m^2) is the collection plate area in the j-th increment. Since the Deutsch Equation is based on the assumption that the migration velocity is constant over the collection area of the precipitator, it is necessary to make the incremental lengths sufficiently small so that the electric field at the plate and the charge accumulated by a given particle size remain essentially constant over the increment.

The collection fraction (fractional efficiency) η_i for a given particle size over the entire length of the precipitator is determined from:

$$\eta_i = \frac{\sum_j \eta_{i,j} N_{i,j}}{N_{i,1}} , \qquad (9)$$

where $N_{i,j}$ is the number of particles of the i-th particle size per cubic meter of gas entering the j-th increment. The quantity $N_{i,j}$ can be written in the form:

$$N_{i,j} = N_{i,j-1} e^{-W_{i,j-1} A_{j-1}/Q} , \qquad (10)$$

where $N_{i,1} = N_{i,0}$, the number of particles of the i-th particle size per cubic meter of gas in the inlet size distribution.

The overall mass collection efficiency η for the entire polydisperse aerosol is obtained from:

$$\eta = \sum_i \eta_i P_i , \qquad (11)$$

where P_i is the percentage by mass of the i-th particle size in the inlet size distribution.

Methods for Representing Non-Ideal Effects

In the preceding section, a basis for calculating ideal collection efficiencies has been developed. This section will discuss the non-idealities which exist in full-scale electrostatic precipitators and describe calculational procedures for estimating the effects on predicted collection efficiencies. The factors of major importance are: (1) gas velocity distribution, (2) gas sneakage, and (3) particle reentrainment due to rapping.

Non-idealities will reduce the collection efficiency that may be achieved for a precipitator operating with a given specific collecting area. Since the model is structured around the Deutsch Equation for individual particle sizes, it is convenient to represent the effect of the non-idealities in the model as correction factors which apply to the exponential argument of the Deutsch Equation. In the subsequent discussions, these correction factors will be used as divisors for the theoretical migration velocities. The resulting "apparent" migration velocities are empirical quantities only and should not be thought of as an actual reduction in the migration velocity in the region of space adjacent to the collecting electrode.

Although it is widely known that a poor velocity distribution gives a lower than anticipated efficiency, it is difficult to apply a numerical description for gas flow quality. White[8] discusses non-uniform gas flow and suggests corrective actions. Preszler and Lajos[9] assign a figure-of-merit based upon the relative kinetic energy of the actual velocity distribution compared to the kinetic energy of a uniform velocity. This figure-of-merit will be a measure of how difficult it may be to rectify the velocity distribution but not necessarily a measure of how much the precipitator performance would be degraded.

It is possible to develop an approach to estimating the degradation of performance due to a non-uniform velocity distribution based upon the velocity distribution, the ideal collection efficiencies, and the Deutsch Equation.[1,2] It will be assumed that the Deutsch Equation applies to each particle size with a known migration velocity and that the specific collection area and size of the precipitator are fixed. A constant k can be determined from the Deutsch Equation such that:

$$k = \frac{A_p W}{A_1} = u_a \ln\left(\frac{1}{1-\eta}\right) \quad , \qquad (12)$$

where A_1 is the total inlet cross-sectional area (m^2) and u_a is the average inlet velocity (m/sec). Now, the mean corrected penetration p for a given particle size can be written as:

$$p = \frac{1}{Nu_a} \sum_{i=1}^{N} u_i \, e^{-\frac{k}{u_i}} \quad , \qquad (13)$$

where N is the number of points in a velocity traverse and u_i are the point values of velocity (m/sec).

For any practical velocity distribution and efficiency, the mean penetration obtained by summation over the velocity traverse will be higher than the calculated penetration based on an average velocity. If an apparent migration velocity for a given particle size is computed based upon the mean penetration and the Deutsch Equation, the result will be a value lower than the value used for calculation of the single point values of penetration. The ratio of the original migration velocity to the reduced migration velocity is a numerical measure of the performance degradation caused by a non-uniform velocity distribution. An expression for this ratio may be obtained by setting the penetration based on the average velocity equal to the corrected penetration obtained from a summation of the point values of penetration, and solving for the required correction factor, which will be a divisor for the migration velocity.

The correction factor "F" may be obtained from:

$$\exp\left(-\frac{k}{F}\cdot\frac{1}{u_a}\right) = \frac{1}{Nu_a}\sum_{i=1}^{N} u_i \exp\left(-k/u_i\right) = p \quad . \qquad (14)$$

Therefore,

$$F = -\frac{k}{u_a(\ln p)} \quad . \qquad (15)$$

Whether the quantity F correlates reasonably well with statistical measures of velocity non-uniformity is yet to be established. A limited number of traverse calculations seem to indicate a correlation between the factor F and the normalized standard deviation of the velocity traverse. Figure 1 shows F as a function of the ideal efficiency for several values of gas velocity standard deviation. These curves were obtained by computer evaluation of Equation 15, and the data on which the calculations are based were obtained from Preszler and Lajos.[9] The standard deviations have been normalized to represent a fraction of the mean. The overlapping of the curves for standard deviations of 1.01 and 0.98 indicates that the standard deviation alone does not completely determine the relationship between F and collection efficiency.

The data in Figure 1 were used to obtain the following empirical relationship between F, the normalized standard deviation of the gas velocity distribution, and the ideal collection predicted for the particle size under consideration:

$$F = 1 + 0.766\, \eta \sigma_g^{1.786} + 0.0755\, \sigma_g \ln \left(\frac{1}{1-\eta}\right) \quad , \qquad (16)$$

where

$$\sigma_g = \frac{\sqrt{\frac{1}{N} \sum_{i=1}^{N} (u_a - u_i)^2}}{u_a} \quad . \qquad (17)$$

This relationship is based on a pilot plant study, and should be regarded as an estimating technique only. If it is desirable to simulate the performance of a particular precipitator, the preferred procedure would be to obtain the relationship between F, η and σ_g for the conditions to be simulated from a velocity traverse at the entrance to the unit.

Gas sneakage occurs when gas by-passes the electrified areas of an electrostatic precipitator by flowing through the hoppers or through the high voltage insulation space. Sneakage can be reduced by frequent baffles which force the gas to return to the main gas passages between the collection plates. If there were no baffles, the percent sneakage would establish the minimum possible penetration because it would be the percent volume having zero collection efficiency. With baffles, the sneakage re-mixes with part of the main flow and then re-bypasses in the next unbaffled area. The limiting penetration due to sneakage will therefore depend on the amount of sneakage gas per section, the degree of re-mixing, and the number of sections.

If the simplifying assumption is made that perfect mixing occurs following each baffled section, an expression for the effect of gas sneakage may be derived. Let:

S = fractional amount of gas sneakage per section,

η = collection fraction of a given size particle obtained with no sneakage for total collection area,

η_j = collection fraction per section of a given particle size = $1 - (1 - \eta)^{1/N_s}$,

N_s = number of baffled sections, and

p_j = penetration from section j.

Then the penetration from section one is given by:

$$p_1 = S + (1 - \eta_j)\ (1 - S)$$

and from section two,

$$\begin{aligned} p_2 &= Sp_1 + (1 - \eta_j)\ (1 - S)p_1 \\ &= p_1\ [S + (1 - \eta_j)\ (1 - S)] \\ &= [S + (1 - \eta_j)\ (1 - S)]^2 \end{aligned}$$

and from section N_s (the last section),

$$\begin{aligned} p_{N_s} &= [S + (1 - \eta_j)\ (1 - S)]^{N_s} \\ &= [S + (1 - S)\ (1 - \eta\ )^{1/N_s}]^{N_s} \quad . \end{aligned} \tag{18}$$

Figure 2 shows a plot of the degradation of efficiency from 99.9% design efficiency versus percent sneakage with number of baffled sections as a parameter. For high efficiencies, the number of baffled sections should be at least four and the amount of sneakage should be held to a low percentage. With a high percentage of sneakage, even a large number of baffled sections fails to help significantly.

A gas bypass or sneakage factor B can be defined in the form of a divisor for the effective, or length averaged, migration velocity in the exponential argument of the Deutsch Equation. The factor B is obtained by taking the ratio of the effective migration velocity w_e under ideal conditions to the apparent value of the effective migration velocity w_e' under conditions of gas sneakage so that:

$$B = \frac{w_e}{w_e'} = \frac{\ln(1-\eta)}{\ln p_{N_s}} = \frac{\ln\ (1-\eta)}{N_s\ \ln\ [S + (1-S)\ (1-\eta)^{1/N_s}]} \quad . \tag{19}$$

Figure 3 shows a plot of the factor B versus sneakage for a family of ideal efficiency curves for five baffled sections. Similar curves can easily be constructed for different numbers of sections.

The foregoing estimation of the effects of sneakage is a simplification in that the sneakage air passing the baffles will not necessarily mix perfectly with the main gas flow, and the flow pattern of the air in the bypass zone will not be uniform and constant. The formula is derived to help in designing and analyzing precipitators by establishing the order of magnitude of the problem. Considerable experimental data will be required to evaluate the method and establish numerical values of actual sneakage rates.

Rapping reentrainment is defined as the amount of material that is recaptured by the gas stream after being knocked from the collection plates by rapping or vibration. With perfect rapping, the sheet of collected material would not reentrain, but would migrate down the collection plate in a stick-slip mode, sticking by the electrical holding forces and slipping when released by the rapping forces. However, the rapping forces are necessarily large to overcome adhesion forces, and much of the material is released into the gas stream as sheets, agglomerates, and individual particles. Most of the material is recharged and recollected at a later stage in the precipitator.

Based upon the simplifying assumptions that a fixed fraction of the collected material of a given particle size is reentrained, and that the fraction does not vary with length through the precipitator, an expression can be derived identical in form to that obtained for gas sneakage:[1,2]

$$P_R = [R + (1-R)\ (1-\eta)^{1/N_R}]^{N_R} \tag{20}$$

where

P_R = penetration corrected for reentrainment

R = fraction of material reentrain

N_R = number of stages over which reentrainment is assumed to occur

η = collection fraction of a given particle size obtained with no reentrainment.

Since Equations (18) and (20) are of the same form, the effect of reentrainment can be expected to be similar to the effect of sneakage, provided that a constant fraction of the material is always reentrained. It is doubtful that such a condition exists, since precipitators frequently use different rapping programs on different sections, agglomeration occurs during collection, and different holding forces exist in different sections. However, until sufficient data on rapping losses PER SECTION as a function of particle size can be accumulated, the relationship may be used to estimate the effect of rapping reentrainment on precipitator performance.

Figure 4 shows the effect on resultant efficiency for a given size particle of various degrees of reentrainment for a four-section precipitator with the indicated values of no-reentrainment efficiency.

Since reentrainment and sneakage effects are estimated with identical mathematical expressions, a combined correction factor B' is used in the mathematical model. From input values of the fraction of material assumed to be lost by reentrainment and sneakage, and the number of stages over which losses are assumed to occur, B' is determined from the ideal collection fraction for each particle size.

In summary, the mathematical model takes into account the non-ideal effects of non-uniform gas velocity distribution, gas sneakage, and rapping reentrainment by reducing the ideally calculated migration velocities w_e by the correction factors F and B'. An "apparent" migration velocity w_e' is determined for each particle size from:

$$w_e' = \frac{w_e}{F \cdot B'} \quad . \tag{21}$$

Using w_e', the corrected fractional collection efficiencies are calculated.

EXPERIMENTAL LABORATORY MEASUREMENTS AND COMPARISON WITH MODEL PREDICTIONS

Laboratory-Scale Precipitator

A laboratory-scale precipitator was constructed for the purpose of studying collection of fine particulate under idealized conditions. The test unit has an electrode irrigation capability and provisions for humidifying flue gas produced from a gas-fired burner. For the experiments described here, however, the carrier gas was ambient air, and the particulate source was an atomizer which produces a polydisperse aerosol of dioctyl phthalate (DOP). Electrode irrigation and humidification were not used, and the collected oil droplets drained from the collection electrodes by gravity.

Figure 5 is a schematic drawing of the laboratory precipitator assembly. Wire-plate geometry is employed, with plate-to-plate spacing and wire-to-wire spacing of 12.7 cm (5 in.). Discharge electrodes consist of 0.269 cm (0.106 in.) diameter wires. The collecting plate area is 2.32 m^2 (25 ft^2) which results in a specific collecting area of 35.0 $m^2/(m^3/sec)$ (178 $ft^2/1000$ cfm) at a gas velocity of 1.37 m/sec (4.5 ft/sec). The precipitator is divided into four length sections separated by baffles to limit bypassage, or sneakage, through the hoppers and the region above the collecting electrode. Three power supplies are used, the first two of which each energize a 76.2 cm (30 in.) length section. The third power supply is connected to the last two sections.

Gas bypassage was estimated by measuring gas velocity with a thermal anemometer in the hoppers and in the region above the collection electrodes. The estimated amount of bypassage, expressed as a fraction of the total gas flow through the unit, was between 8% and 10% for the two values of gas flow reported here. Gas flow uniformity was determined by conducting a traverse with the thermal anemometer at the precipitator inlet. The standard deviation of the velocity distribution, expressed as a percentage of the average velocity, was 11.2% for an average gas velocity of 1.09 m/sec.

Measurement Technique

Fractional efficiency measurements were conducted using a five-stage Brink impactor at the precipitator inlet and outlet. Aluminum foil substrates were used to facilitate weighing of the small masses collected on the individual stages. The mass determinations were made with a Cahn Electrobalance. For the experiments conducted at the lower gas velocity, a sampling time of six hours was required at the precipitator outlet to obtain weighable quantities on the impactor substrates. Six and four point traverses were conducted with the impactor at the inlet and outlet, respectively, in order to insure that a representative size distribution was obtained at each location.

Since the particulate exiting from an electrostatic precipitator will have an electrical charge, it was necessary to employ a charge neutralizer at the inlet of the sampling line for all measurements at the precipitator outlet. Measurements with an optical particle counter indicated that, if metal substrates were used in the impactor, the charge acquired by the particulate substantially influenced the size distribution obtained with the impactor. Since the impactor was calibrated with an uncharged aerosol, the use of the charge neutralizer minimized sizing errors due to electrostatic attraction.

Table 1 gives a typical set of inlet and outlet data obtained from duplicate experiments at the indicated conditions. Good reproducibility was obtained for all stages at the inlet, and for stages 2 through 5 at the outlet. Isokinetic sampling was not employed, since the effect of anisokinetic conditions on particles with diameters less than 2.0 μm is expected to be negligible. The effect of changing the gas flow rate through the impactor on the indicated size distribution was checked by reducing the flow rate from 56.6 cm^3/sec to 28.3 cm^3/sec. The higher flow rate, which was used for all of the fractional efficiency determinations, resulted in a gas velocity of 0.76 m/sec at the sample point. Figure 6 shows the inlet size distributions from Table 1 and the distribution obtained at the same location with the reduced impactor flow rate. These data indicate that the same size distribution is obtained at either flow rate. The higher flow rate is more desirable because of the smaller cut points and the increased amount of gas that may be sampled per unit time.

Experimental and Theoretical Results

Fractional efficiency measurements were performed with the Brink impactor and DOP aerosol at current densities of 26.9, 53.8, and 107.5 nA/cm^2 at a gas velocity of 1.37 m/sec, and at current densities of 26.9 and 53.8 nA/cm^2 at a gas velocity of 0.72 m/sec. The results obtained from these experiments, expressed as collection efficiency as a function of particle size, are presented in Figures 7, 8, and 9. Computed collection efficiencies obtained with the mathematical model are also shown. In view of the difficulties in making these types of measurements, the agreement between measured and calculated efficiencies shown in Figure 7 is considered good. Figures 8 and 9 suggest that the agreement obtained between computed and measured results decreases with decreasing gas velocity and current density. The assumption of 8% gas sneakage over four stages improves agreement between computed and measured results at a gas velocity of 0.72 m/sec. However, the data obtained at this lower gas velocity with 26.9 nA/cm^2 are considerably below the computer projections. Possible causes of this lack of agreement are unmodelled effects such as non-uniform current density and electric field, and particle concentration gradients in the inter-electrode space.

Another series of experiments is being conducted at the present time in which fractional efficiencies are being measured for different wire-to-plate spacings and wire sizes. Figure 10 shows experimental and theoretical fractional collection efficiencies and migration velocities for a plate-to-plate spacing of 19.05 cm (3.75 in), wire radius of 0.1588 cm (0.125 in), and gas velocity of 1.46 m/sec (4.79 ft/sec). These results indicate that the model predictions give good agreement with experimental data for current densities of 26.9 na/cm^2 or less. However, for current densities greater than 26.9 na/cm^2 it appears that the agreement worsens and, in fact, the ideally calculated fractional efficiency curve lies below the measured curve.

The disagreement in Figure 10 between theory and experiment for current densities greater than 26.9 na/cm^2 occurs at a plate spacing and gas velocity which are not too different from those encountered in the field. Since space charge effects are not significant in these experiments, it might be suspected that particle concentration gradients exist under certain conditions and in these cases the collection mechanism in the model is not strictly applicable.

Measurements have been made at the outlet of the laboratory scale precipitator under several operating conditions to determine whether or not concentration gradients existed. A "Climet" optical particle counter was used to count 1.0, 0.5, and 0.3 μm diameter particle sizes at several positions located between the collecting plate and the wire. The measurements at the 12.7 cm plate spacing under all conditions presented here showed no build-up of particles near the collecting electrode. These results are consistent with the fact that the model yields good agreement with experiment under all conditions at the 12.7 cm plate spacing. The measurements at the 19.05 cm plate spacing showed definite concentration gradients existing for all three particle sizes under several conditions. In these cases, the concentration was greatest near the collecting plate and decreased towards the discharge electrode. This type of behavior might be expected due to the directed coulomb motion of a particle which is superimposed on the turbulent motion of the gas stream. The ratios of the concentration near the collecting plate to that near the wire for 1.0, 0.5 and 0.3 μm diameter particles were 1.7, 1.3, and 1.3, respectively, at approximately 43.0 na/cm^2. The causes for such gradients, the conditions under which they would exist, and their effect on predicted collection efficiency of fine particles need further investigation.

FIELD MEASUREMENTS AND COMPARISON WITH MODEL PREDICTIONS

In this section the predictions of the mathematical model are compared with experimental field data for several differing precipitators. These precipitators were involved in the collection of coal flyash under various operating conditions. Also, "typical" performance curves are generated using the model for "cold" precipitators with operating current densities from 5 to 40 na/cm^2 and for a "hot" precipitator operating at a current density of 30 na/cm^2.

Figure 11 gives a comparison of fractional effective migration velocities obtained from the model with those obtained from inertially-determined fractional efficiency measurements on a coal-fired power boiler. This "cold" full-scale precipitator has a low specific collection area (SCA) and moderate current density. The particulate under collection had a mass median diameter (MMD) of approximately 25 μm and a low value of resistivity. In addition to the ideal predictions, Figure 11 shows the effect of correcting for a gas velocity standard deviation (σ_g) of 0.25. A comparison of the w_e values in Figure 11 indicates that the theoretical predictions are low for the particle diameters below 2.0 μm diameter and high for larger particles. Due to limitations in the measurement technique, no data are available for particle diameters larger than 5.0 μm. Figure 12 shows the fractional efficiency data from which the w_e values of Figure 11 were calculated. The overall mass efficiency predicted by the model was 99.8% as compared to a measured value of 99.6%.

Figures 13 through 18 give results predicted by the model in terms of overall mass efficiency as a function of SCA for current densities of 5, 10, 15, 20, and 40 na/cm^2 at 150°C, and for a current density of 30 na/cm^2 at 370°C. The relationships presented in these figures are based on a 22.86 cm (9 in) plate spacing, and secondary volt-amp curves and particle size distribution (MMD = 25 μm) considered to be typical for a coal-fired power boiler. Also given are test results obtained under conditions approximating the electrical conditions represented by the given values of current density. A comparison of the limited amount of applicable test data with the computed results indicates that the ideally-predicted overall mass efficiencies are generally higher than those obtained from the field measurements. The measurements were taken with sampling techaniques which insured that essentially all of the particulate mass larger than 0.3 μm diameter was captured by the sampling device.

The use of the empirical correction factors reduces the computed values of overall mass collection efficiency to the range of values obtained from the field measurements. Reentrainment and sneakage effects were combined by assuming that the indicated fractional losses per stage occurred over four effective stages. Note that high efficiency (greater than 99.5%) precipitators operating on the low temperature side of the air heater fall reasonably close to the computed line obtained with $\sigma_g = 0.25$ and $S = 0.1$. A gas velocity distribution with a standard deviation of 0.25 is generally considered to be a good distribution for a full-scale unit. The computed results show that a poor distribution ($\sigma_g = 0.5$) seriously degrades performance. The detrimental effects of sneakage and reentrainment are also indicated in the decrease of computed performance resulting from variations in the parameter S.

Figures 19 and 20 show results obtained from the analysis of a "cold" full-scale precipitator having a large SCA [560 $ft^2/1000$ ft^3/min)] and low current density (12 na/cm^2). The experimental results were obtained from inlet and outlet size distribution measurements obtained using inertial, optical, and electrical sizing techniques. The precipitator is preceded by a mechanical collector so that the inlet size distribution consisted of finer particles with an MMD of approximately 3 μm. The fly ash under collection had a high resistivity resulting in limitations on the current density. The measurements were performed with no rapping so that reentrainment should be minimized. The ideally predicted fractional efficiency and migration velocity curves lie below the experimental data for particle sizes in the 0.06-2.5 μm range. Over this size range the theoretical curve has the same shape as the experimental data with the minimum occurring at about 0.4 μm in both cases. The overall mass efficiency predicted by the model is 99.7% as compared to the measured value of 99.9+%.

Figures 21 and 22 show results obtained from a study of a "cold" pilot precipitator operating at various SCA's and current densities ranging from 40 to 7 na/ft^2. The experimental results were obtained from inlet and outlet measurements with mass trains using in-stack filters. Particle size distribution measurements made at the pilot precipitator inlet with inertial impactors indicated that a typical fly ash distribution was obtained and that for particle diameters below about 5 μm, the cumulative mass loadings obtained in the main duct were within the range of those obtained at the pilot plant inlet.

Both Figures 21 and 22 indicate that in order to match the test data with the predicted model results at comparable current densities, it is necessary to either assume a poor gas velocity distribution ($\sigma = 0.50$) or relatively large losses due to reentrainment and gas bypassage. The bypassage and reentrainment losses are assumed to occur over three effective stages, since the pilot unit consisted of three sections. Gas flow measurements with a thermal aneomometer indicated that at this installation gas flow quality was poor ($\sigma = 0.46$ or greater), and that gas sneakage was estimated to be no smaller than 8% and no greater than 30% of the total gas flow. Resistivity data from in-situ measurements and voltage current curves from the pilot unit power supplies indicated that the current densities of 11 and 32 na/cm^2 could be sustained without experiencing electrical breakdown of the dust layer deposited on the collection electrodes under the test conditions.

Figure 23 shows results obtained from the analysis of a "hot" full-scale precipitator having an SCA of 430 $ft^2/(1000$ $ft^3/min)$ and current density of

35 na/cm^2. The experimental results were obtained from inlet and outlet size distribution measurements obtained using inertial, optical, and diffusional sizing techniques. For this installation, the fractional efficiency curve obtained from the measurements is lower than the theoretical curve obtained from the mathematical model. Although the average electrical operating conditions for this unit are used in the model, some of the electrical readings for individual sets were characterized by anomalously low current and voltage levels, suggesting the possibility of electrode misalignment. This is a contributing factor in the disagreement between theoretical and measured results since the average electrical conditions will not adequately represent the effects of low voltages and current in individual sets on collection efficiency. The theoretically predicted overall mass efficiency was 99.9% at the indicated condition, and the average measured collection efficiency was 99.3%. If a gas velocity distribution standard deviation of 0.25 is assumed, sneakage and reentrainment losses of 10 to 20% over three stages are required to reconcile the computed and measured mass efficiency.

CONCLUSIONS

Comparisons of theoretically calculated collection efficiencies with laboratory data obtained under essentially idealized conditions indicate that the mathematical model gives good agreement with experiments for particle sizes from 0.55 to 2.5 μm diameter over a wide range of current densities at gas velocities of 0.72 and 1.37 m/sec with a 12.7 cm plate spacing. For a 19.05 cm plate spacing and gas velocity of 1.46 m/sec, the model predicts ideal fractional efficiency curves which lie below the measured data for current densities greater than 25 na/cm^2. This discrepancy may at the present be attributed to unmodelled concentration gradients which were measured for 0.3, 0.5, and 1.0 μm diameter particles under these conditions.

Comparisons of model predictions with field data obtained from full-scale and pilot precipitators indicate that the theory generally gives a good prediction of overall mass collection efficiency although, in certain cases, the fine portion of the fractional efficiency curve lies below the measured data. The laboratory measurements obtained to date indicate that the cases in which the model underpredicts fine particle collection may be due to concentration gradients which exist under certain conditions.

Considering the broad range of laboratory and field data with which the theory has been compared, it can be concluded that the mathematical model provides a basis for indicating performance trends caused by changes in specific collecting area, electrical conditions, and particle size distribution, provided that back corona does not exist. Current density, applied voltage, and the particle size distribution are the most important variables in the calculation of overall mass collection efficiency for a given specific collection area. The theoretical calculation of ideal overall collection efficiency of polydisperse particulate in an electrostatic precipitator generally gives results higher than those obtained from performance measurements on coal-fired power boilers. Corrections to the idealized or theoretical collection efficiency to estimate the effects of non-uniform gas flow, rapping reentrainment, and gas bypassing the electrified sections reduce the overall values of calculated efficiency to the range of values obtained from field measurements. These calculations suggest that the theoretical model may be used as a basis for quantifying performance under field conditions if sufficient data on the major non-idealities become available.

ACKNOWLEDGMENTS

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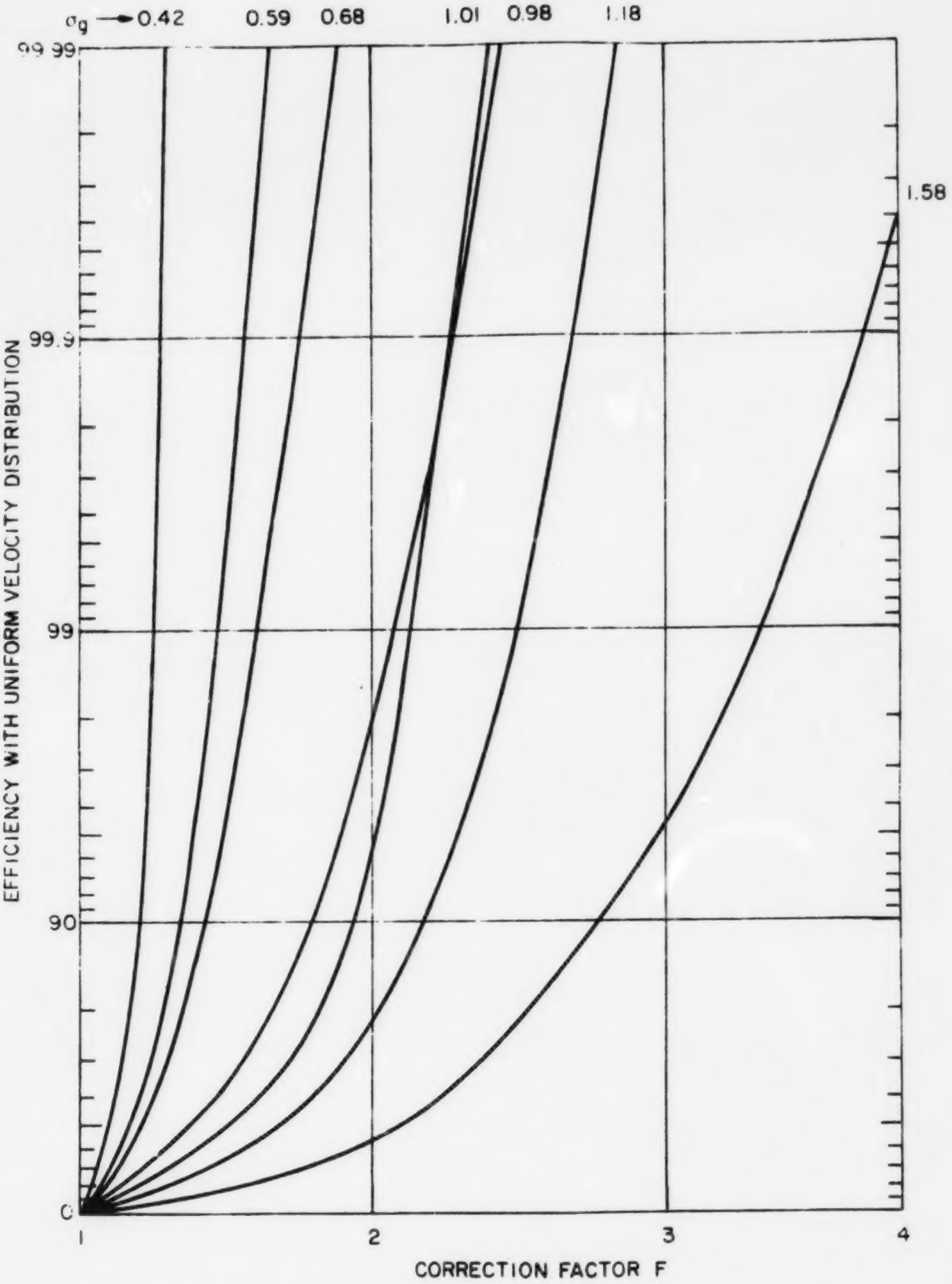

Figure 1. "F" as a function of ideal efficiency and gas flow standard deviation.

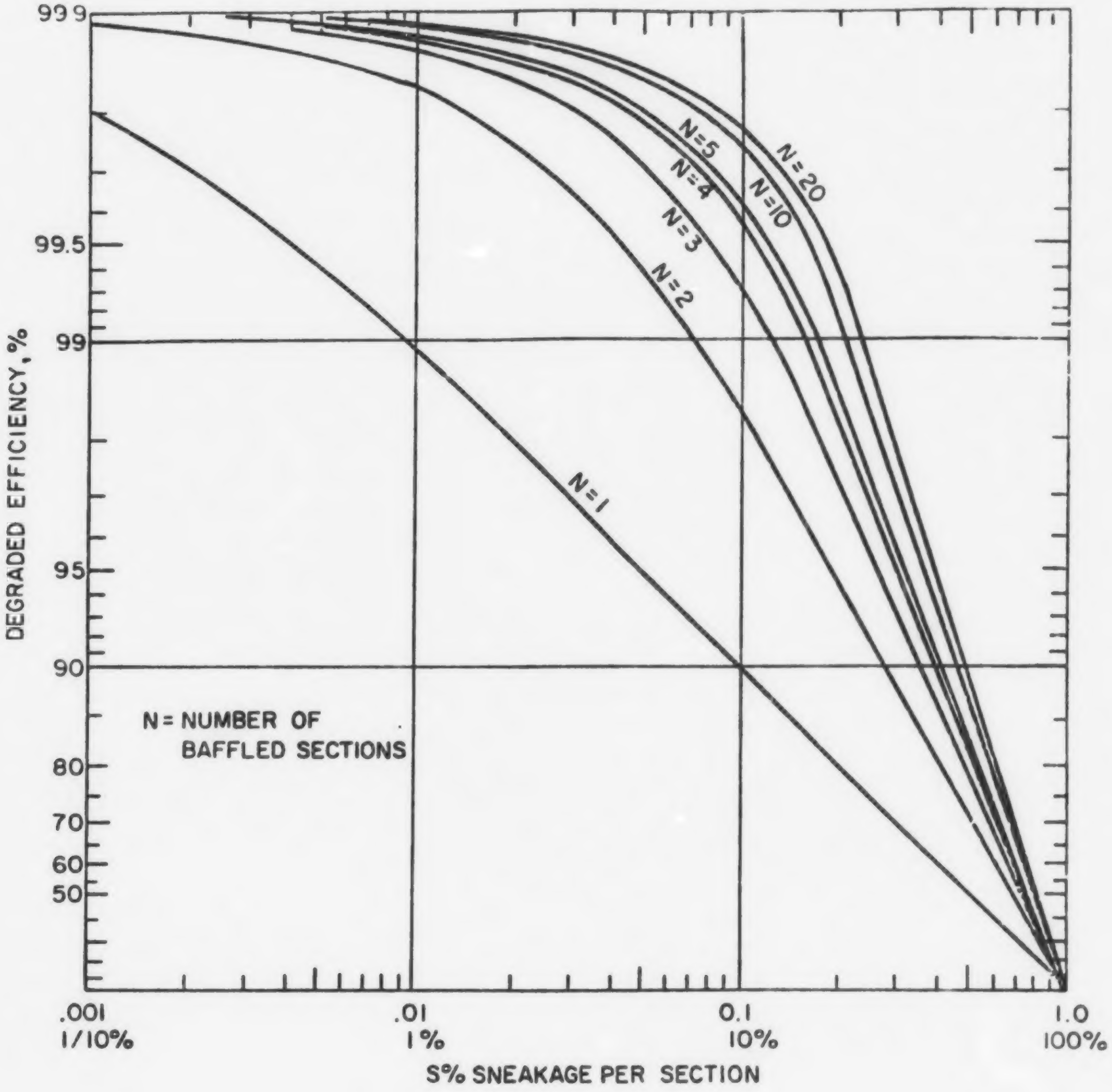

Figure 2. Degradation from 99.9% efficiency with sneakage.

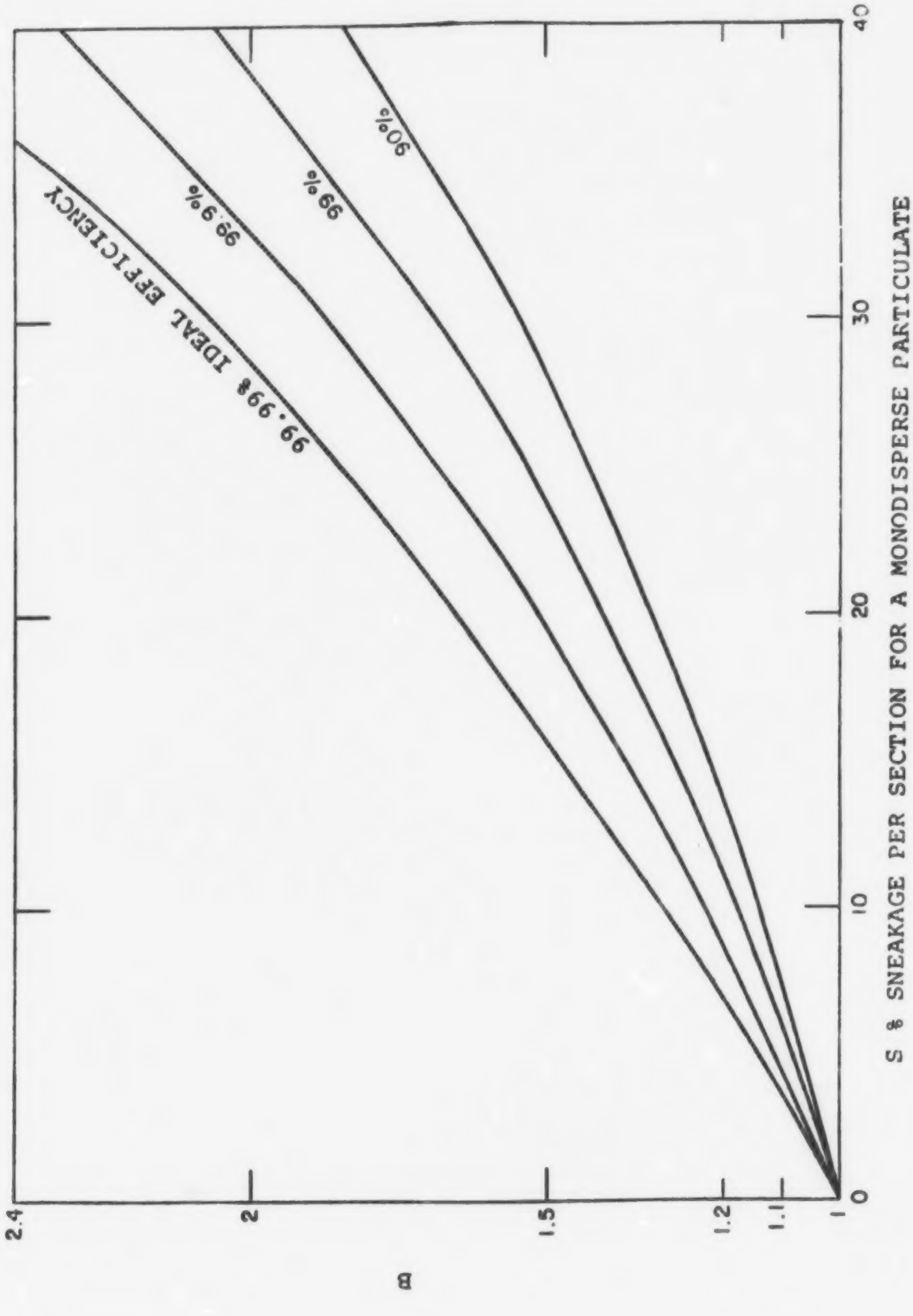

Figure 3. Correction factor for by-pass sneakage when $N_s = 5$.

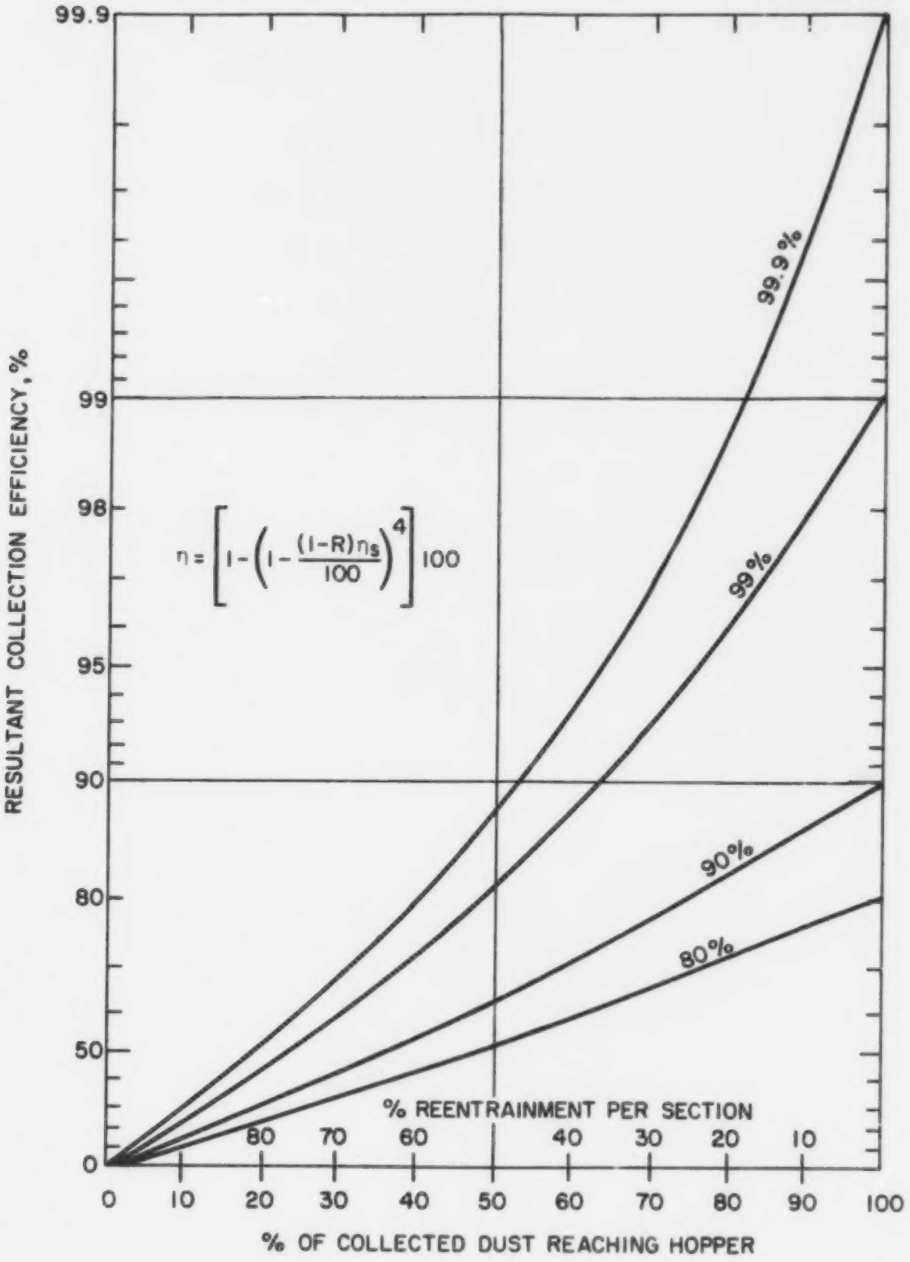

Figure 4. Effect of reentrainment on the efficiency of a four-section precipitator designed for a no reentrainment efficiency as indicated for a monodisperse particulate.

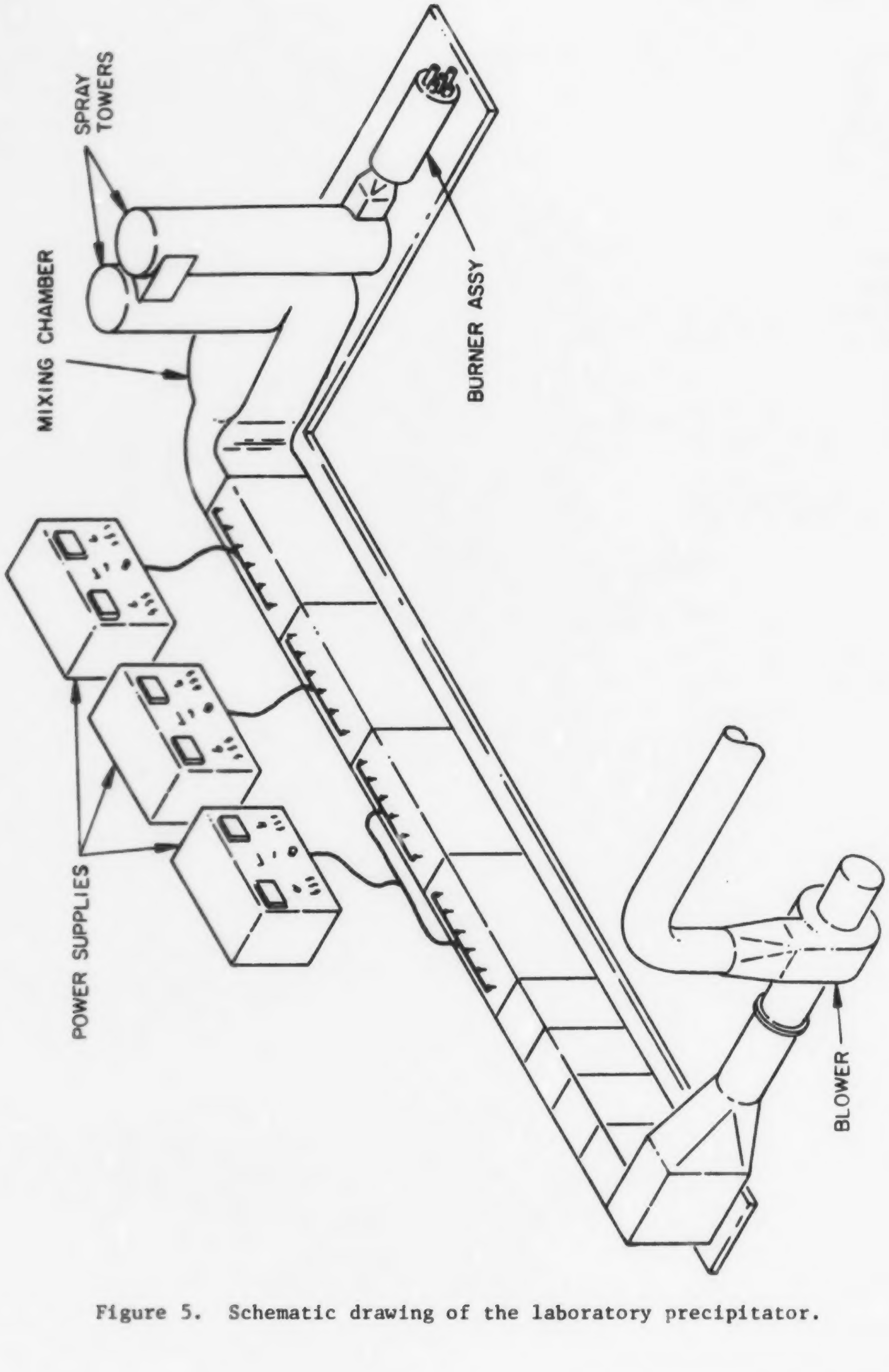

Figure 5. Schematic drawing of the laboratory precipitator.

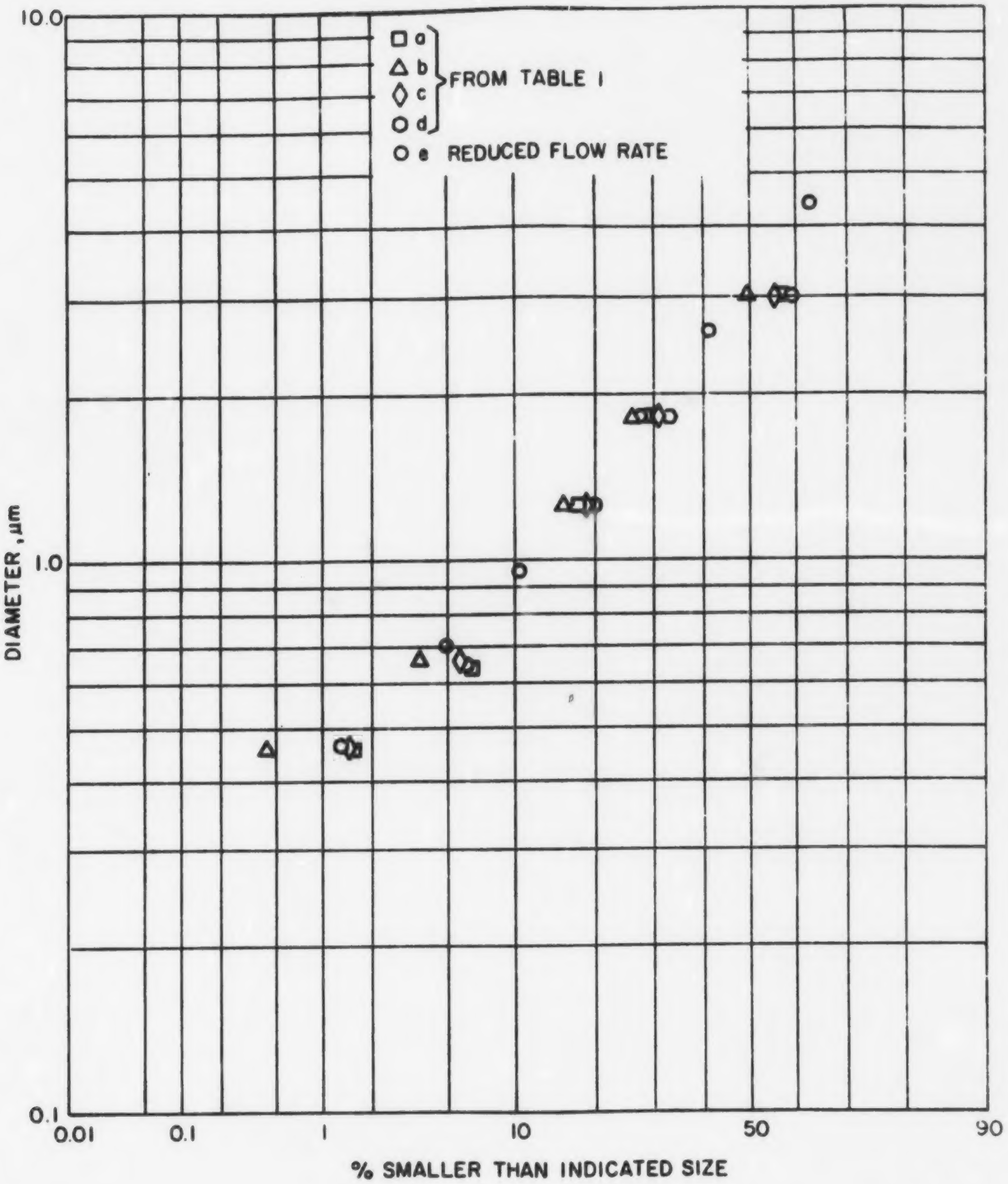

Figure 6. Inlet particle size distributions from Brink data.

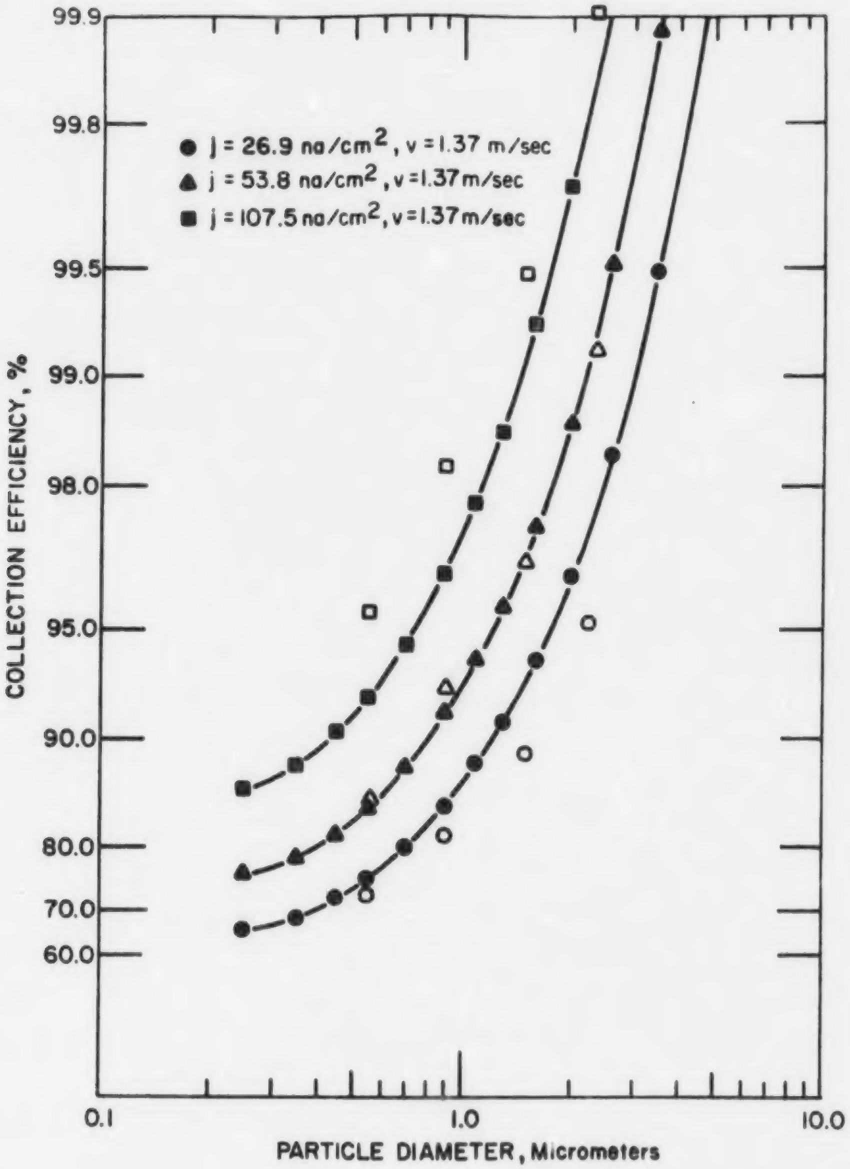

Figure 7. Experimentally measured and ideally calculated fractional collection efficiencies in the laboratory precipitator.

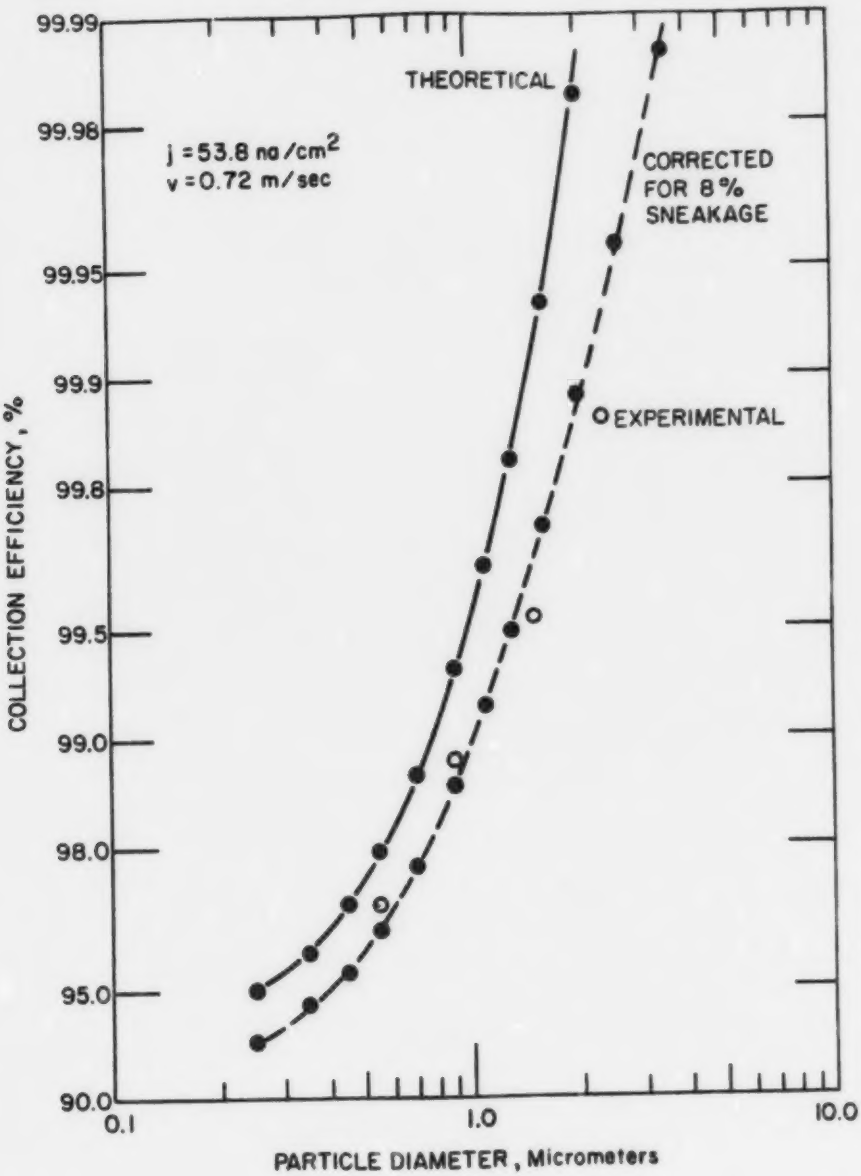

Figure 8. Experimentally measured, ideally calculated, and sneakage-corrected fractional collection efficiencies for a current density of 53.8 nA/cm^2 and a gas velocity of 0.72 m/sec in the laboratory precipitator.

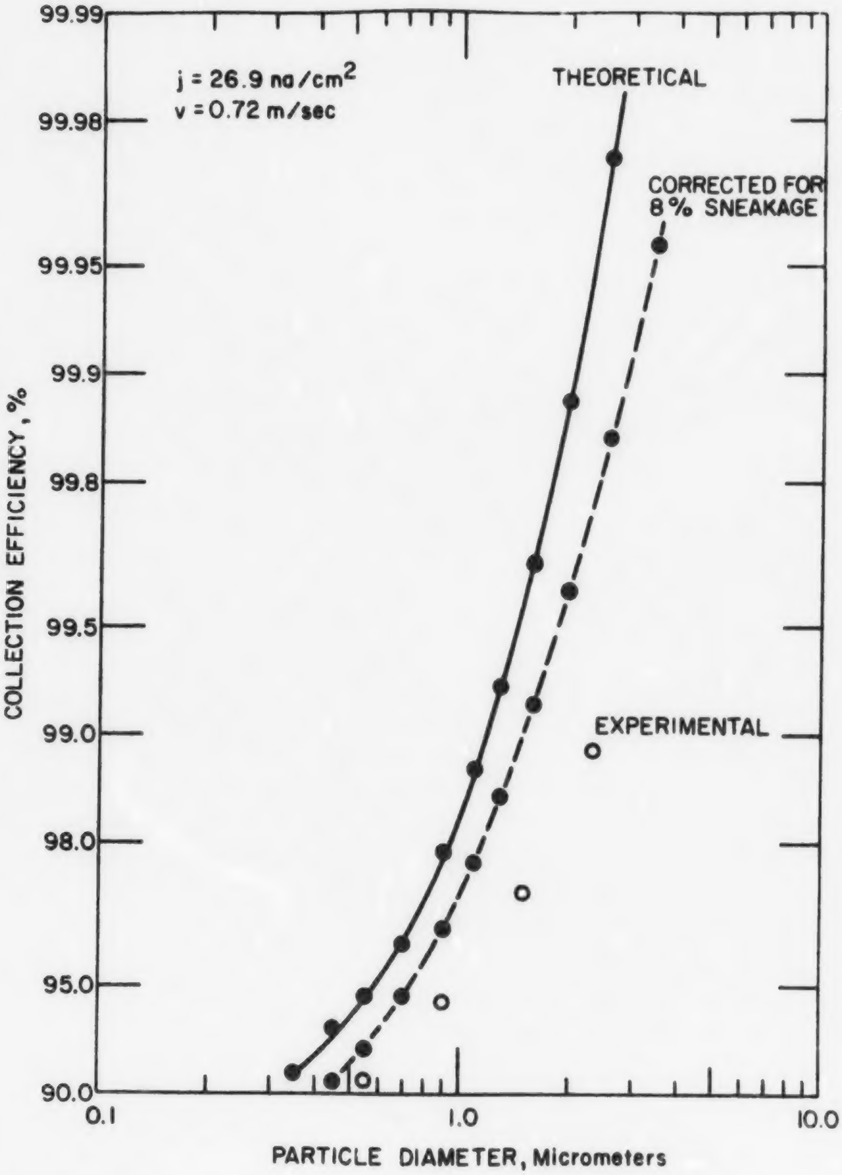

Figure 9. Experimentally measured, ideally calculated, and sneakage-corrected fractional collection efficiencies for a current density of 26.9 nA/cm^2 and a gas velocity of 0.72 m/sec in the laboratory precipitator.

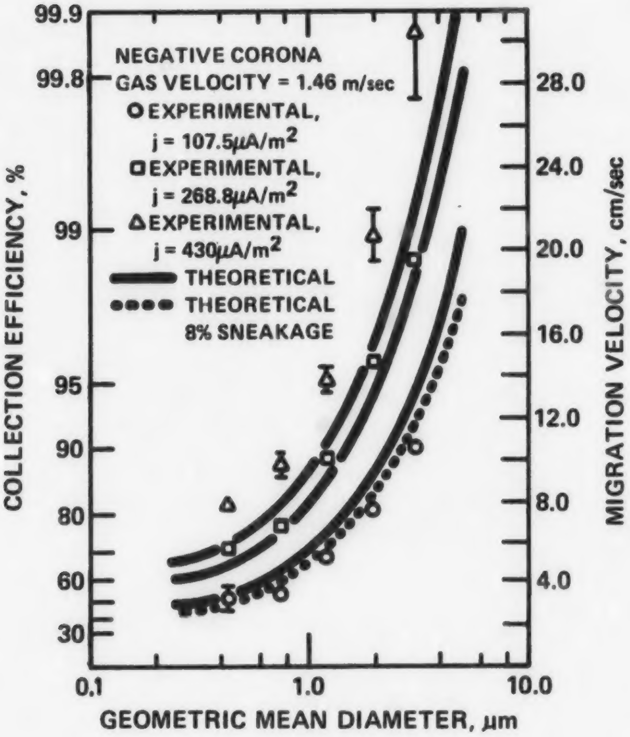

Figure 10. Experimental and theoretical data for a wire of radius 1.488×10^{-3}m.

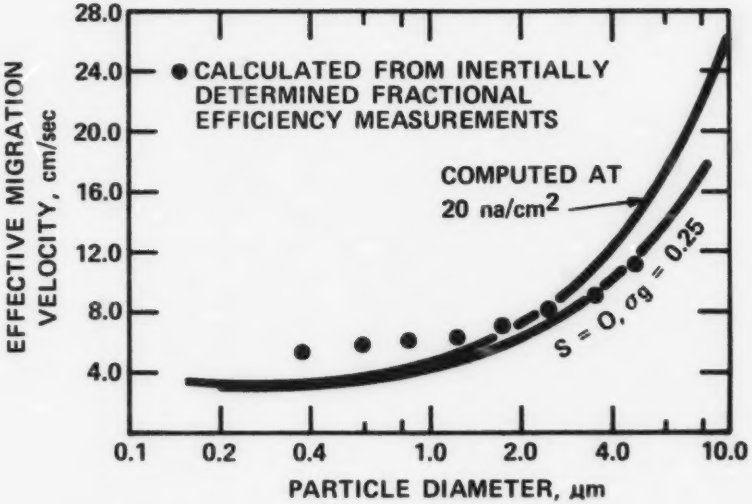

Figure 11. Effective migration velocities for a full-scale precipitator on a coal-fired boiler (SCA = 55.7 $m^2/(m^3/sec)$.

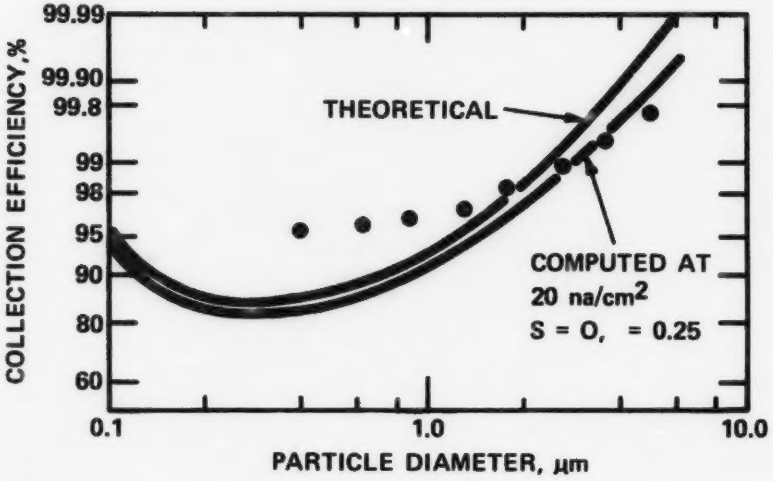

Figure 12. Fractional collection efficiencies for a full-scale precipitator on a coal-fired power boiler.

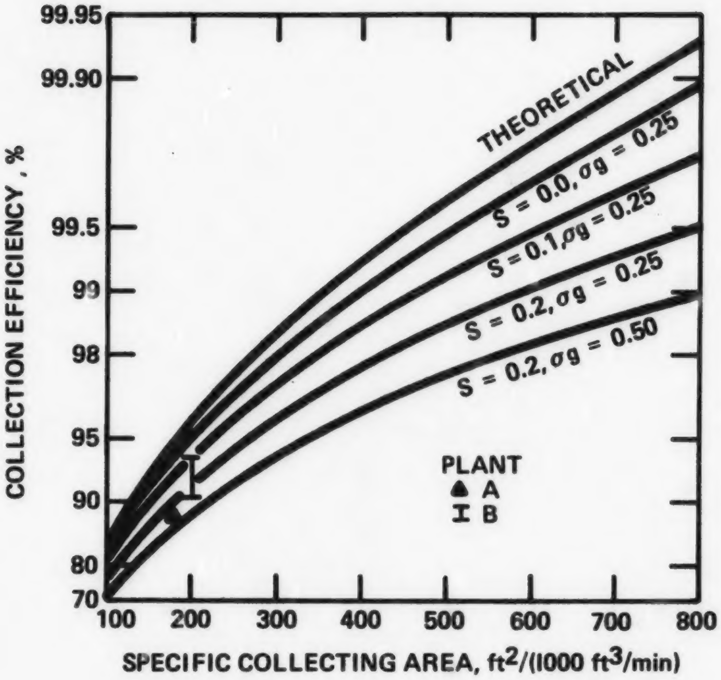

Figure 13. Computed performance curves at 5 nA/cm^2.

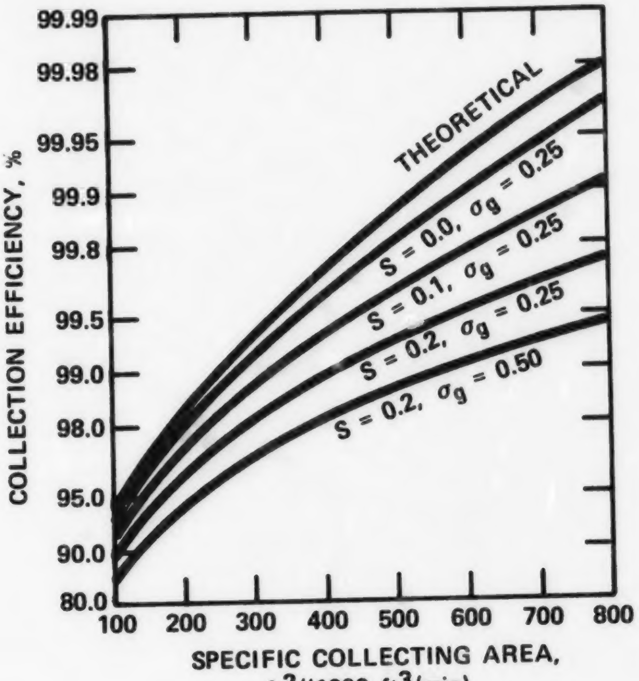

Figure 14. Computed performance curves at 10 nA/cm^2.

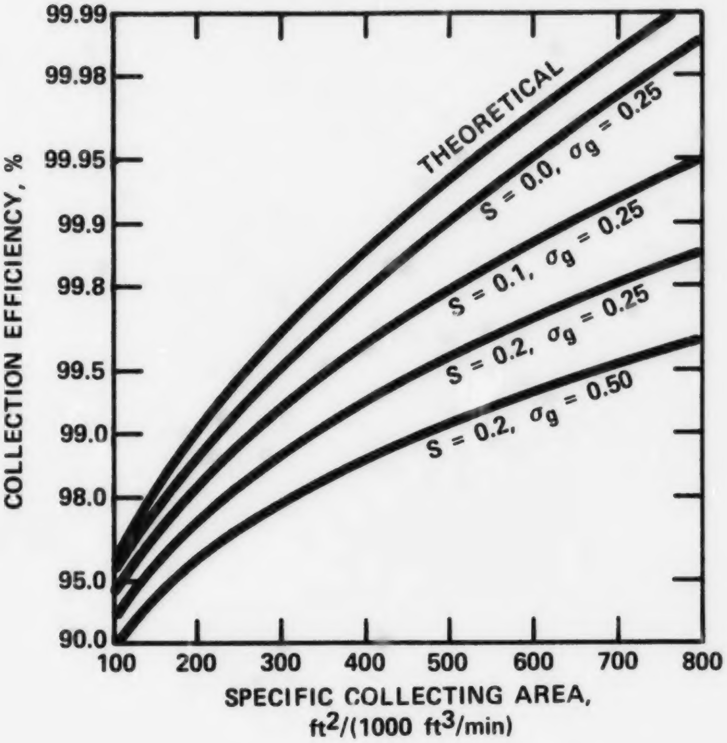

Figure 15. Computed performance curves at 15 nA/cm^2.

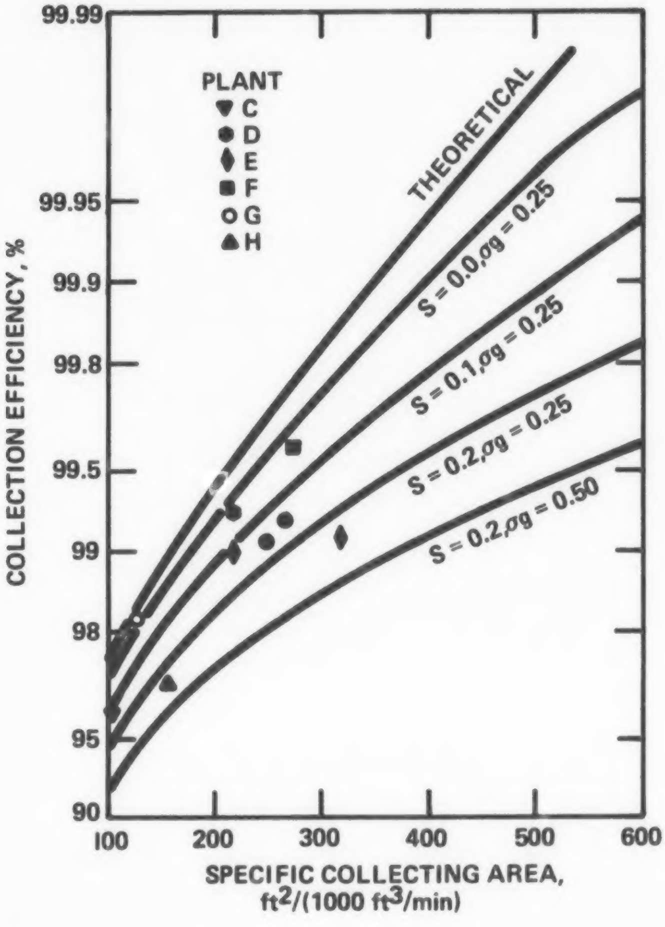

Figure 16. Computed performance curves at 20 nA/cm^2.

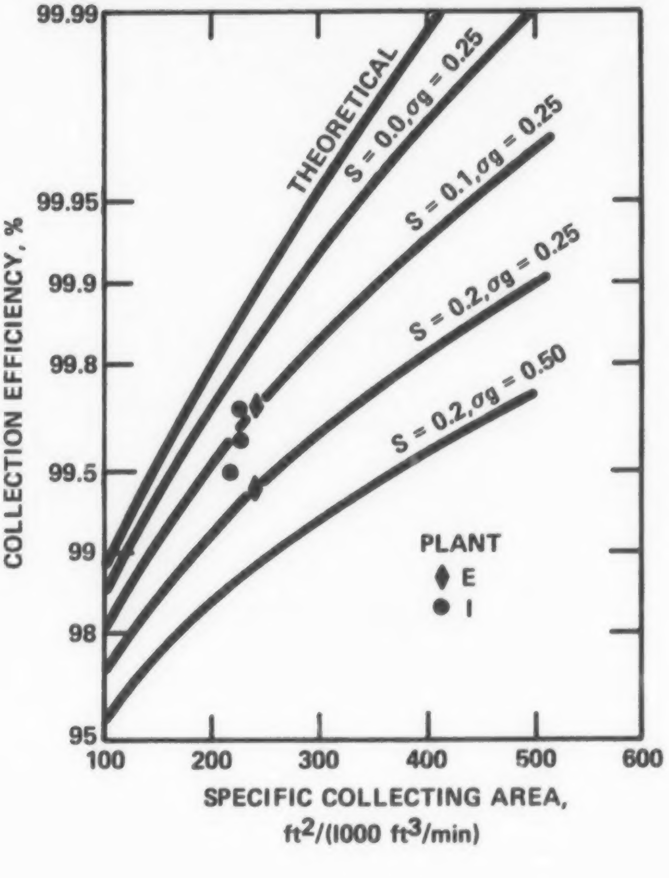

Figure 17. Computed performance curves at 40 nA/cm^2.

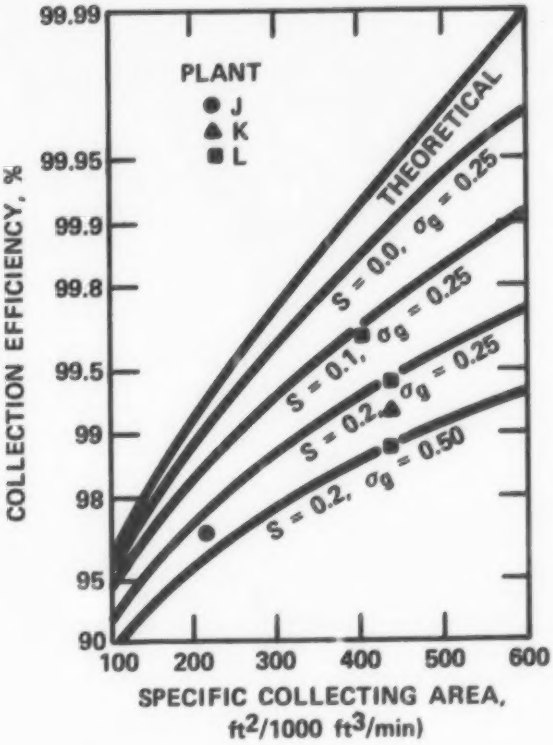

Figure 18. Computed performance curves for "hot" precipitator.

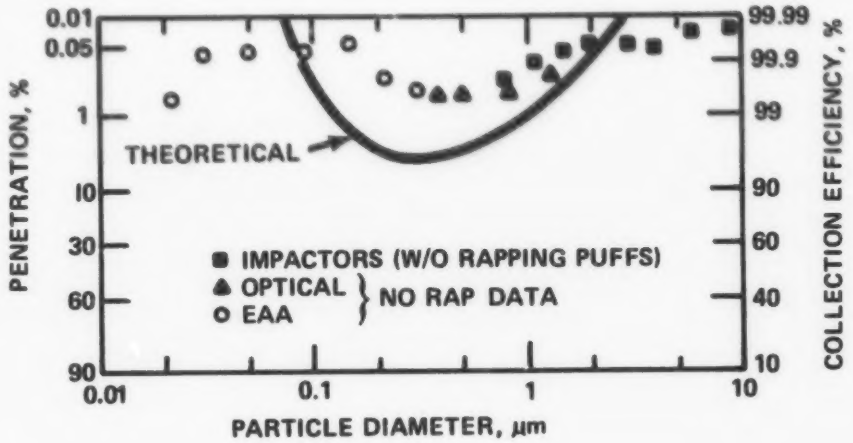

Figure 19. Measured and theoretical fractional efficiency.

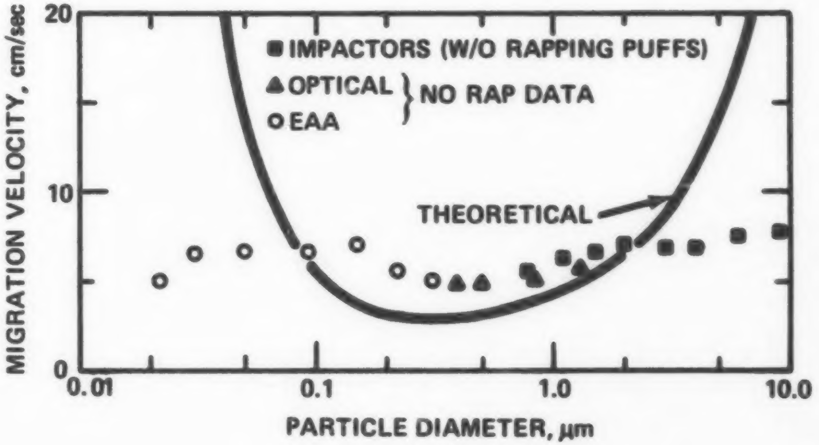

Figure 20. Effective migration velocities vs particle diameter.

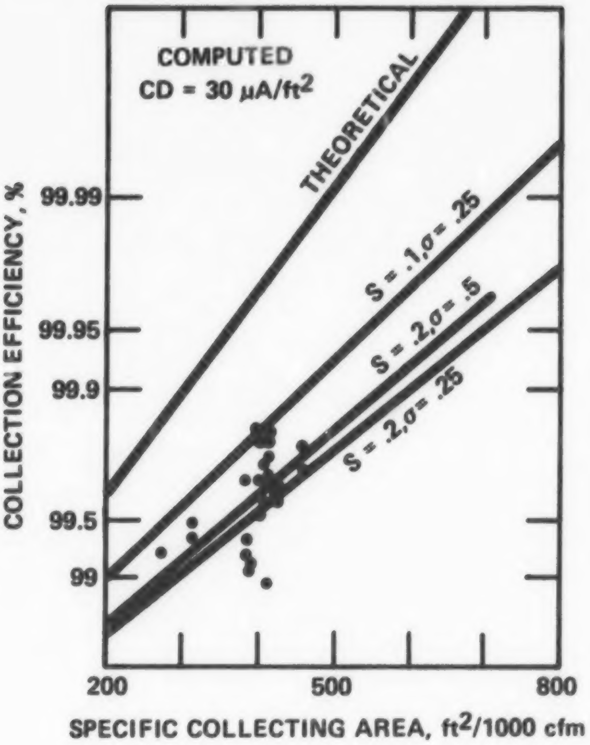

Figure 21. Computed and measured data from pilot precipitator.

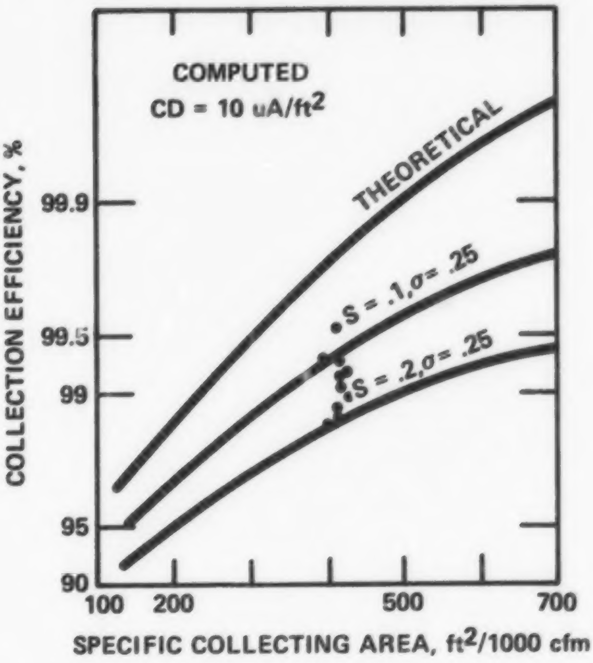

Figure 22. Computed and measured data from pilot precipitator.

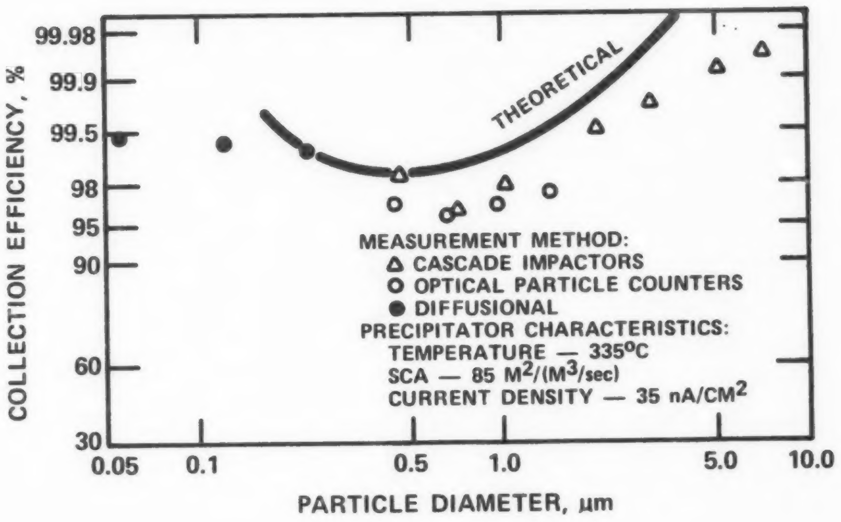

Figure 23. Comparison of measured and computed fractional collection efficiencies.

Table 1

TYPICAL INLET AND OUTLET PARTICLE SIZE DATA FOR WET ESP EXPERIMENTS
(DOP Sprayer, Gas Velocity = 1.37 m/sec, Current Density = 55 na/cm^2)

Test No.		Mass Loading, mg/am^3									
		Inlet					Outlet			Geometric Mean Diameter, µm	Collection Efficiency, %
Stage	Lower Size Limit, µm	2ai	2bi	2ci	2di	Average	2ao	2bo	Average		
1	3.04	26.341	28.956	24.596	24.088	25.995	0.0414	~0	0.207	-	-
2	1.80	18.659	13.526	14.128	15.226	15.385	0.1494	0.1103	0.1299	2.34	99.16
3	1.24	7.249	6.285	7.238	8.226	7.250	0.2575	0.2115	0.2345	1.49	96.77
4	0.65	7.448	6.984	6.743	7.419	7.148	0.5724	0.4782	0.5253	0.90	92.65
5	0.46	2.954	1.775	2.514	2.654	2.474	0.3862	0.3770	0.3816	0.55	84.58

IMPROVED PRECIPITATOR TECHNOLOGY BY PILOT PLANT TESTING AND EVALUATION OF COAL CORE-CORES

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IMPROVED PRECIPITATOR TECHNOLOGY
BY PILOT PLANT TESTING AND EVALUATION OF COAL BORE-CORES

by

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Abstract

When coal from a new mine field is to be used in a proposed pulverized-coal-fired boiler, it is important to be able accurately to size the precipitator needed to collect the flyash. Especially if the unknown coal is from the Western United States the ash resistivity may be high, which increases the size of the precipitator required for a specified collection efficiency. As much information as possible on this point may be derived from laboratory tests of coal samples from an adequate number of bore-cores from the mine field, early in the precipitator planning and design. Small-scale combustors, burning 2 to 75 pounds of pulverized coal per hour, can produce flyash representative of that eventually produced by the planned full-scale boiler. Thus the ash resistivity and particle size distribution may be determined in advance, facilitating more accurate precipitator size selection.

Pilot precipitator tests, properly interpreted, permit a correlation to be evolved between effective migration velocity (a measure of precipitator performance) and the ash resistivity and particle size distribution. Thus the correct precipitator size may be more accurately predicted by a combination of coal bore-core analysis and pilot precipitator testing. Such testing is to be elaborated soon by construction of a new experimental facility at Arapahoe Station, Denver, Colorado.

IMPROVED PRECIPITATOR TECHNOLOGY

BY PILOT PLANT TESTING AND EVALUATION OF COAL BORE-CORES

1. Introduction

When a power company ("Utility") plans to build a new generating plant, to be coal-fired, it is often necessary to open a new mine field to provide the required coal. To minimize air pollution and meet the legal standards for particulate emission, the flyash produced by the coal burning is usually collected from the combustion gases by an electrical precipitator. Such precipitators may allow only $\frac{1}{2}$% of the ash to escape up the stack if it comes from "coal A", but, say, 2%, or four times as much, to escape if it comes from another "coal B".

When the utility asks several precipitator manufacturers to bid, given the gas flow rate, temperature, required collection efficiency, etc, the bids it receives may vary by a factor as great as four, in some instances. It is obviously desirable to reduce this spread to a lower maximum such as, say 2 to 1, or hopefully a maximum such as, say, 1.1 to 1. Technical methods for doing this are the subject of this paper.

2. New Coal Mine Field Survey

The new coal mine field may have an area of several square kilometers or square miles. A survey of the field in the early planning stages for the new power station results in a geological map. This should include a series of isopleths, each connecting points of a chosen overburden depth, for example. Likewise for coal seam depth.

Tests of the coal from each of the dozens of bore-cores for such properties as heat of combustion, for example, may permit one to draw isopleths for this on the field map, also. Ashing part of each of these coal samples, followed by chemical analyses of the resulting ash, may permit one to add isopleths for a "boiler-fouling index", for example.

Since there is some evidence of a relationship between ash (chemical) composition and ash resistivity (to be discussed later), one might further attempt to add approximate isopleths for ash resistivity to be expected from coal mined from various parts of the field.

Finally, as described later, ash resistivity and ash precipitability are related, so one might even attempt rather speculative isopleths for ash precipitability. Examination of these isopleths may permit the precipitator designer more accurately to size the new precipitator to operate satisfactorily, even on ash from the unfavorable parts of the new coal field.

3. Combustors

Since it is the ash that the precipitator collects, not the coal, it is very desirable to find a way to produce an ash sample from the coal that may be available from the new mine field, so that the ash may be tested in the laboratory to estimate how easy or how difficult it may be to precipitate. To do this, it is necessary to burn the coal in such a way that its ash will very closely resemble the ash produced later when that coal is burned in the full-scale completed boiler. The device designed to burn the pulverized coal in the laboratory in this special way is usually called a "combustor".

Several such combustors already exist. As far as we are aware, the first one built primarily to help predict precipitator performance on flyash is the "ACIRL combustor" operated by the Australian Coal Industry Research Laboratories[1]. Some of the results obtained to date with this combustor have been described by Tassicker and Sullivan[2]. Figure 1 shows this ACIRL "micro-combustor" which burns only a few grams of coal per hour.

Figure 2 shows a larger combustor recently built by Exxon Research and Engineering Co. in Baytown, Texas, their chief investigator being E. C. Winegartner. It burns about 5 kilograms of coal per hour. Figure 3 shows a comparable combustor in the Babcock & Wilcox Laboratories in Alliance, Ohio, known as their "Laboratory ashing furnace" or "LAF"; however, we shall call it the "B & W combustor". We understand that their chief investigator, C. L. Wagoner, plans to modify this combustor in four respects: (1) insert a liner into the furnace zone; (2) sectionalize the boiler portion of the equipment; (3) add transmissometer capability; and (4) change the coal feeder so as to obtain coal rates of 1 or 2 kilograms per hour, from the present 2 to 9, while retaining the same volumetric heat input rate.

At this (San Francisco) symposium, Selle and Hess have described some of the recent work[3] with the pulverized coal combustor at the Energy Research Center, U. S. Bureau of Mines, Grand Forks, North Dakota. Figure 4 shows this combustor and the experimental precipitator connected so as to collect the flyash from the coal which is burned at a rate around 35 kilograms per hour. There is also a combustor of comparable size in Australia at the Commonwealth Scientific and Industrial Research Organization, or "CSIRO". Its coal-burning rate is about 25 kilograms per hour, and some of the work with it is mentioned by Tassicker[4].

The Aerotherm Division of Acurex Corporation in Mountain View, California has built and is now operating a "Multi-fuel furnace" rated at 3/4 million kilocalories/hour (3 million BTU/hour). With coal as a fuel, this amounts to about 100 kilograms or 250 pounds per hour. This combustor was designed to burn natural gas, distillate oil, No. 6 oil, or pulverized coal. It was designed originally to study methods of reducing NO_x emission. Those attending the (San Francisco) symposium had an opportunity to inspect this combustor in operation, at Mountain View. It is shown in Figure 5 and was described for the visitors by R. Brown. The future flyash studies are to be conducted by J. Steiner.

To summarize these combustors in order of nominal coal burning rates, we have: (1) ACIRL, a few grams per hour; (2) B & W, 2 kg/hr; (3) Exxon, 5 kg/hr; (4) CSIRO, 25 kg/hr; (5) Grand Forks, 35 kg/hr; and (6) Acurex, 100 kg/hr. There may be others.

4. Combustor Technology

If the combustor is to produce ash like that from the same coal burned in a full-scale boiler, the coal must be pulverized before it is fed into the combustor, much as it is for the boiler. A jet of air, flowing at the proper rate to yield the desired percentage of excess air (above stoichiometric), is usually used to carry the pulverized coal, fed at the desired rate. into the combustion chamber. When the combustor first starts up, there must be provision for igniting the mixed coal dust and air. Some laboratories use a "pilot" butane burner to do this. Others heat the walls of the combustion chamber with electrical heaters to such a high temperature that the air-coal mixture ignites spontaneously. The combustion chamber walls are made of a refractory material capable of containing the burning coal at temperatures between 1100 and 1650°C.

The hot, ash-laden gases from the combustion chamber may then (a) go to a small test precipitator for collecting the ash, or (b) go to a fiber-glass bag for ash collection, the ash to be re-dispersed later in a synthetic gas stream for precipitation, or more usually the bagged ash can be tested for resistivity, particle size distribution, etc, using separate apparatus. In any of these cases, the ash-laden gas from the combustor must be cooled from, say, 1400° to, say, 150° C before it is led to the test precipitator or glass bag. A heat exchanger is required to do this.

While going from the combustor through the cooling exchanger to the test precipitator or glass bag the gas velocity throughout should be kept in the vicinity of 15 to 20 meters per second. Otherwise a large fraction of the ash will settle out before it reaches the precipitator or bag. Thus the diameter of the duct leaving the combustor might be, say, 8 centimeters, where the gas temperature is, say, 1400° C, but the duct should gradually taper down to, say 4 centimeters as the gas enters the precipitator or bag, to compensate for the contraction of the gases as they cool while moving along.

To quote from a B & W report about their "LAF" combustor: "Our experience has indicated that the residence time for combustion in the LAF of about 0.9 second is a good model of typical full-scale values in the range of 0.8 to 1.0 second.

However, a heat rate of about 700,000 kilocalories/meter^3/hour (80,000 BTU/ft^3/hr) in the LAF is required to simulate full-scale ash sintering characteristics produced at firing rates between 90,000 and 130,000 kcal/m^3/hr (11,000 and 15,000 BTU/ft^3/hr). This is at least in part due to the high temperature of the furnace walls in the LAF and the resulting reduction in apparent furnace volume".

The foregoing discussion points up the need for due care in designing and operating a laboratory combustor to produce flyash practically identical to that from a full-scale boiler burning the same coal. Can such a combustor follow the difference observed between ash from a cyclone-fired boiler and ash from other types of boiler burning the same coal?

A few of the combustors, such as the one at Grand Forks, for example, are directly connected to a test precipitator, and can be so operated that the gas from the combustor, after going through the heat exchanger, can be made to enter the precipitator at a chosen temperature such as, say, 150° C if that is the temperature anticipated in the future full-scale precipitator. With such a laboratory system, one may run "coal A", say, and then later, "coal B". If the former yields an "effective migration velocity" (w) of 10 cm/sec, for example, and the latter yields w = 5 cm/sec, this indicates that the final full-scale precipitator should be about twice as big and twice as costly if "coal B" is to be burned as would be necessary if "coal A" were burned, other things (gas flow rate, collection efficiency, gas temperature, etc.) being equal.

5. Correlation Between Pilot Precipitator Migration Velocity (w), Ash Resistivity, and Particle Size Distribution

If the value of w can not be obtained directly by connecting a laboratory precipitator to the combustor as just described, the alternative is to collect the ash in a glass bag and later, using separate equipment, measure the bulk resistivity of the ash, its particle size distribution, etc, and from these data estimate the value of w to be anticipated in the new full-scale precipitator, for that coal and that ash, under the expected conditions such as temperature, etc.

Wolcott[5] was the first to publish quantitative data relating the electrical resistivity of a dust to its precipitability, in 1918. Since then, apparatus and procedures for determining bulk resistivity of dust samples have been described by a number of investigators[6], and official standards, such as the one by APCA[7] have been established. A typical recent resistivity graph showing the variation of resistivity of an Australian flyash with temperature and humidity may be seen in Figure 6 from an earlier reference[2].

On the left or cool side of the peaks in the curves in Figure 6, a typical dust, in bulk, conducts electricity mainly via surface layers on each particle, these adsorbed layers consisting typically of moisture, and in the case of flyash, sulfur trioxide, and other salts. On the right or hot side of the peaks in Figure 6, most of these adsorbed surface layers evaporate and the conduction is mainly via the interior of the dust particles. The volume conductivity increases with rising temperature, as with a typical semi-conductor, but the surface conductivity decreases as the temperature rises because the rise evaporates more and more of the adsorbed conducting surface layer. Thus the resistivity (the inverse of conductivity) reaches a maximum typically around 110 to 150° C, depending upon the humidity and dust particle composition.

In 1955 Sproull[8] published the graph in Figure 7, showing in a general way how the effective migration velocity (w) attainable in precipitating limestone dust varies with its resistivity. Note that a resistivity of 2×10^{12} ohm-cm corresponds to a value of w around $3\frac{1}{2}$ cm/sec, which is very poor. In addition to resistivity, migration velocity is also affected by many other factors, including gas density, viscosity, sparkover voltage, and by particle loading. Despite such ambiguities, Figure 7 presents a general picture of the way the performance of a single-stage (conventional) precipitator passes through a transition from good below 10^{10} ohm-cm to poor above 10^{11} ohm-cm.

Recent resistivity graphs for some different ashes from Western coals, in a paper by Walker[9] are shown in Figure 8. He notes that although these coals range in sulfur content only from 0.17 to 0.91%, at temperatures below 205° C (400° F) the ash resistivity ranges over five orders of magnitude. Thus the sulfur content of these low-sulfur coals does not help one predict their ash resistivities. Table 1, from the same paper by Walker, lists pertinent data from four stations operating on Wyoming sub-bituminous coal, which he found to be more readily interpreted by using the equation

$$F = 1 - e^{-(wf)^r} \qquad (1)$$

where F = collection efficiency (as a decimal, not a percentage), w has already been defined, f = "specific collecting area" = collecting electrode area/gas flow rate, and r is an arbitrary exponent selected to give the best fit with the data. If r = 1, the above equation becomes simply the Deutsch equation, but several precipitation investigators in recent years have found that a better fit with experimental data is often attained if r is assigned values other than 1.

In equation (1), the specific collecting area (often called "S C A"), f, should be expressed in seconds/cm if w is in cm/sec, as it usually is. The folks who use the metric system often put f in seconds/meter, with w in meters/second. Precipitation engineers in the U. S. commonly express f, or "S C A" in square feet per 1000 cfm, as Walker has done in Table 1. Multiply these values by 0.1968 to convert to seconds/meter.

Table 1.

Summary of Precipitator Performance Tests on Western Coals
(From paper by A. B. Walker, expanded)

Case No.	Test No.	Precipitator temperature °F	°C	Efficiency %	S C A ft^2/1000 acfm.	S C A sec/meter	In situ resistivity ohm-cm	Drift vel, w cm/sec
1	1	330	166	99.2	253.4	50.0	1.98×10^8	9.66
	2	330	166	99.1	253.4	50.0	1.98×10^8	9.42
	3	300	149	99.5	664.1	131.	6.5×10^9	4.03
	4	300	149	98.8	664.1	131.	6.5×10^9	3.38
	5	300	149	99.9	664.1	131.	6.5×10^9	5.26
2	1	330	166	94.1	184.3	36.3	9.7×10^9	7.80
	2	330	166	91.6	204.9	40.2	7.9×10^9	6.15
	3	330	166	94.8	204.4	40.2	7.4×10^9	7.35
3	1	885	474	99.2	235.8	46.4	---	10.4
	2	885	474	98.5	235.8	46.4	---	9.08
4	1	345	174	87.0	99.1	19.5	1.6×10^{12}	10.45
	2	309	154	89.3	111.4	21.95	7.3×10^{11}	10.15
	3	300	149	89.1	120.4	23.7	2.6×10^{11}	9.35
	4	292	144	78.9	103.9	20.5	3.4×10^{11}	7.60
	5	301	150	82.9	107.3	21.2	1.9×10^{11}	8.35
	6	305	152	81.3	109.5	21.5	2.3×10^{11}	7.80
	7	293	145	88.8	125.2	24.6	1.6×10^{11}	8.90

Using the Deutsch equation (equation (1) with $r = 1$), we have calculated the drift velocity, w, from the values of efficiency (F) and SCA (f) given in Table 1. These calculated values of w are in the last column, next to the ash resistivity values in the adjacent column. If resistivity were the only factor to be considered, w should always increase as resistivity decreases, but the table shows that it does not. Among other factors to be considered is the ash particle size distribution. The Deutsch equation is based on the assumption that the particles are of uniform size; they are not. A prime reason for introducing the arbitrary exponent r into equation (1) was to adjust the equation to fit the experimental data better in this respect[10].

Particle size distribution is commonly measured with an instrument like the "Bahco analyzer". For flyash, the size distribution is close to "log normal", which means that ordinarily it can be approximated by a straight line on the proper type of logarithmic graph paper, as in Figure 9, from Walker's paper[9].

Figure 10 by Hewitt[11] shows how the drift velocity, w, reaches a minimum at a particle size ("diameter") near 0.3 micron, both theoretically and experimentally, for positive corona. Industrial precipitators use negative corona, but the corresponding curves closely resemble those in Figure 10, as seen in Figure 11, from a recent paper by McCain, Gooch, and Smith[12]. The performance of an industrial precipitator is not as sensitive to particle size as Figure 10 indicates, because considerable agglomeration occurs in all precipitators. Examination of the particle size distributions of the ash collected in the "A", "B", and "C" fields (in series) in a precipitator reveals a surprising uniformity.

In general, other things being equal (and they rarely are), a larger precipitator will be needed for unusually fine ash than for ash of ordinary coarseness.

6. Ash Resistivity Related to Its Chemical Composition

Investigators are devoting substantial effort to finding a correlation between the resistivity of flyash and its chemical composition, but more work is needed in this area to establish a sufficient confidence level. Elements such as sodium, potassium, and iron in the coal are also found in its ash, and these appear to some extent to take over the role of sulfur (in high-sulfur coals) in reducing the surface resistivity of the ash particles when sulfur and SO_x becomes scarce, as in most Western coals. Selle and Hess[13] and Bickelhaupt[14] are among the pioneers in studying this effect, and preliminary indications are that more sodium, potassium, and iron in the coal and its ash tends to reduce its surface resistivity. For a given required collection efficiency, this resistivity reduction tends to increase w, thus allowing a reduction in f or SCA, and hence in precipitator size and cost.

These elements do not occur in the ash as such, but rather, in such forms as Na_2O, K_2O, and Fe_2O_3. These compounds, when present in the ash, obviously got there from the coal when it burned, so they must have been in the coal. Na_2O sublimes at 1275° C or 2327° F and Fe_2O_3 melts at 1565° C or 2849° F, so these compounds simply pass through the combustion process and emerge in the ash, despite the high furnace temperature. Predicting ash (surface) resistivity from elemental analysis of the coal (and its ash) from bore cores thus provides a complementary approach to that from producing flyash from a combustor.

7. Correlation-Regression Equations

From data contained in the paper by Tassicker and Sullivan[2], one may write a regression equation such as

$$w = a_o + a_1 \ln_e f + a_2 \log_{10} \rho + a_3 \rho + a_4 (T/100) + a_5 (T/100)^2 + a_6 (1/d) + a_7 z + a_8 \log_{10} z + a_9 (T/100)^3 \qquad (2)$$

where w = effective migration velocity, f = specific collecting area, as already defined, ρ = resistivity of the ash, T = temperature in $^\circ$ C, d = particle size in microns, and z = ash content of the coal, in percent.

Using pilot precipitator test data from 161 tests, with coal (ash) from 5 mines, 3 seams, with ash contents (z) from 16.9 to 22.1%, f values for the precipitators from 30 to 120 seconds/meter (9 to 36 seconds/foot), average particle sizes ranging from 9.6 to 14.4 microns, resistivities from 5×10^8 to 5×10^{12} ohm-meters (5×10^{10} to 5×10^{14} ohm-cm), and gas temperatures from 85 to 350° C, the following results were obtained from application of equation (2):

Coefficient	Value	Reduction in variance	Comments
a_0	+1.06146	19.6%	Coefficient is a kind of "dump".
a_1	−0.25149		Steady coefficient in all tests.
a_2	−0.02077	33.2%	Most significant reduction in variance.
a_3	$+4.0436 \times 10^{-15}$		
a_4	−0.12856	20.7%	Steady coefficient in all tests.
a_5	+0.03037		
a_6	+0.09565	1 %	Slight statistical significance.
a_7	+0.02099	7 %	Low statistical significance.
a_8	−0.74618		
a_9			Significant above 250° C in Figure 13.

Multiple correlation coefficient $R = 0.899$

Standard error $SE = 0.0085$ meters/second.

The above data, as well as Figures 12 and 13 are from a paper by Tassicker[15]. The meaning of terms such as "Variance", "Correlation coefficient", and "Standard error" may be found in appropriate textbooks[16].

Figure 12 illustrates how some of the data from the above-mentioned 161 tests fit the regression equation, as a function of f or SCA. Figure 13 shows the fit for temperature. Similar analytical approaches have been published by Coy and Frisch[17].

8. "Exact" Mathematical Analytical Models

Instead of assigning r a value differing from 1 in equation (1), one can stay with the basic Deutsch equation ($r = 1$) and apply it repeatedly to dust particles in narrow size ranges such as, say, 0.2 to o.4 micron, then 0.3 to 0.4 micron, etc, up to, for example, the fraction larger than 50 microns. Then the results can be added properly to arrive at a weighted average value of f, say, if appropriate values of w have been chosen for each size fraction. This is too tedious unless one uses a computer, as Gooch and Francis[18] have done. A simplified outline of their computer program is shown in Figure 14. This program, to date, has been applied to flyash only.

Inspection of Figure 14 does not reveal how such information as expected gas temperature and moisture content, expected resistivity of the ash (at that temperature and humidity), etc, can be fed into the program. It was probably designed for use in cases where no data are available from bore cores from a new mine field, so that an experimental approach could be used, as already outlined.

In fact, Reynolds, Theodore, and Marino[19], who have developed a mathematical model along similar lines, state in their paper: "A model for calculating collection efficiencies—one based solely on physical principles and, as a result, one requiring no prior information from pilot plant or field testing—has been employed to generate performance curves for twelve industries presently using electrostatic precipitators". The twelve industries are: (1) Flyash (electric power), (2) Municipal incinerators, (3) Gypsum, (4) Cement, (5) Pulp (and paper) salt cake, (6) Sintering, (7) Steel open hearth, (8) Blast furnaces, (9) Coke oven detarring, (10) Sulfuric acid mist, (11) Phosphorus mist, and (12) Petroleum cracking.

Several precipitator manufacturers are reported to have such computer programs either in use or under development. Just how successful and dependable these may be is not known. Coy and Frisch[17] state that they use "various approaches, including statistical ones, to improve the reliability of the sizing under conditions of minimal information".

9. EPRI Pilot Precipitator Experimental Station to be Built Arapahoe Station

Experimenting with full-size precipitators in an effort to improve their performance, or to try out new techniques oe unorthodox designs in an effort to advance the technology, is prohibitively expensive because of their great size and cost. Then too, power plant owners and operators do not wish to vary the operation of thei plant (load factor, excess air, etc.) just to accommodate the precipitator testing. A relatively small pilot precipitator is therefore used for such work, for the gas flow rate through it may be varied by merely adjusting a damper, or a change in the electrode system may be made quickly and relatively easily without disturbing the station in any way. There is also a great saving in cost and manpower, the work proceeds faster, and there are other reasons.

After the desired tests are finished, the pilot precipitator is ordinarily removed from the plant, later to be shipped to some other plant for other tests. Test results obtained at the second plant may not be compared accurately with those at the first plant because a different coal may be used, or the boilers may differ in type and size. Thus there is an advantage in establishing a permanent pilot precipitator experimental station at a selected power plant, so that many ideas and designs may be tested and compared with each other, and experiments such as those outlined in this paper may be conducted routinely.

For these reasons, EPRI has made arrangements to establish a permanent pilot precipitator experimental station at Arapahoe Station on the outskirts of Denver, this station being owned and operated by the Public Service Co. of Colorado. The plant burns Western coal routinely. Construction of the new experimental precipitation facility is expected to begin this year (1976).

It is anticipated that new ideas will be tested, such as, for example, improved ways of charging the ash or dust, ways to agglomerate the ash more effectively than it is now in a conventional precipitator, etc. Moreover, if the above-described combustor tests with small coal samples from bore cores are insufficient in some cases, perhaps a coal shipment from the new mine could be burned and the ash tested at Arapahoe.

Figure 15 is a schematic plan of the new facility, showing two pilot precipitators nominally rated at 8,500 meters^3/ hour (5,000 cfm) each, and two others nominally rated at 60,000 m^3/hr (35,000 cfm) each. These are ducted so that the two small ones can be operated either in series or in parallel, on gas from either the hot or the cold side of the air preheater. Likewise the two large pilot precipitators can be operated either in series or in parallel, but only on gas from the cold side of the heater. The laboratory building is to be located in the center, as shown. It will contain all of the controls and instruments, which will be much more elaborate than those normally accompanying the usual movable pilot precipitator.

10. Conclusions

The task of the precipitator designer is to select a precipitator for a given installation, which will be as small as possible, as inexpensive as possible, and which will perform dependably at the required collection efficiency when operating at the specified full-load gas flow rate, burning the specified coal in boilers of the chosen type and size.

If only a few kilograms of the coal is available, the designing task is difficult, as this paper has shown. Even if the coal is available in large quantities, the task is still complex. The developments described here represent significant progress along these lines, and when some or all of these new techniques are operable, the designing of new full-scale precipitators will be greatly facilitated. No longer should a power company receive bids for a new precipitator differing by a factor such as 2 or more. A factor such as 1.1 should be expected eventually. Certainly more work along these lines is needed.

Finally, the new test facility at Arapahoe Station should lead to significant basic developments and technical "break-throughs" in electrical precipitation, in the years to come.

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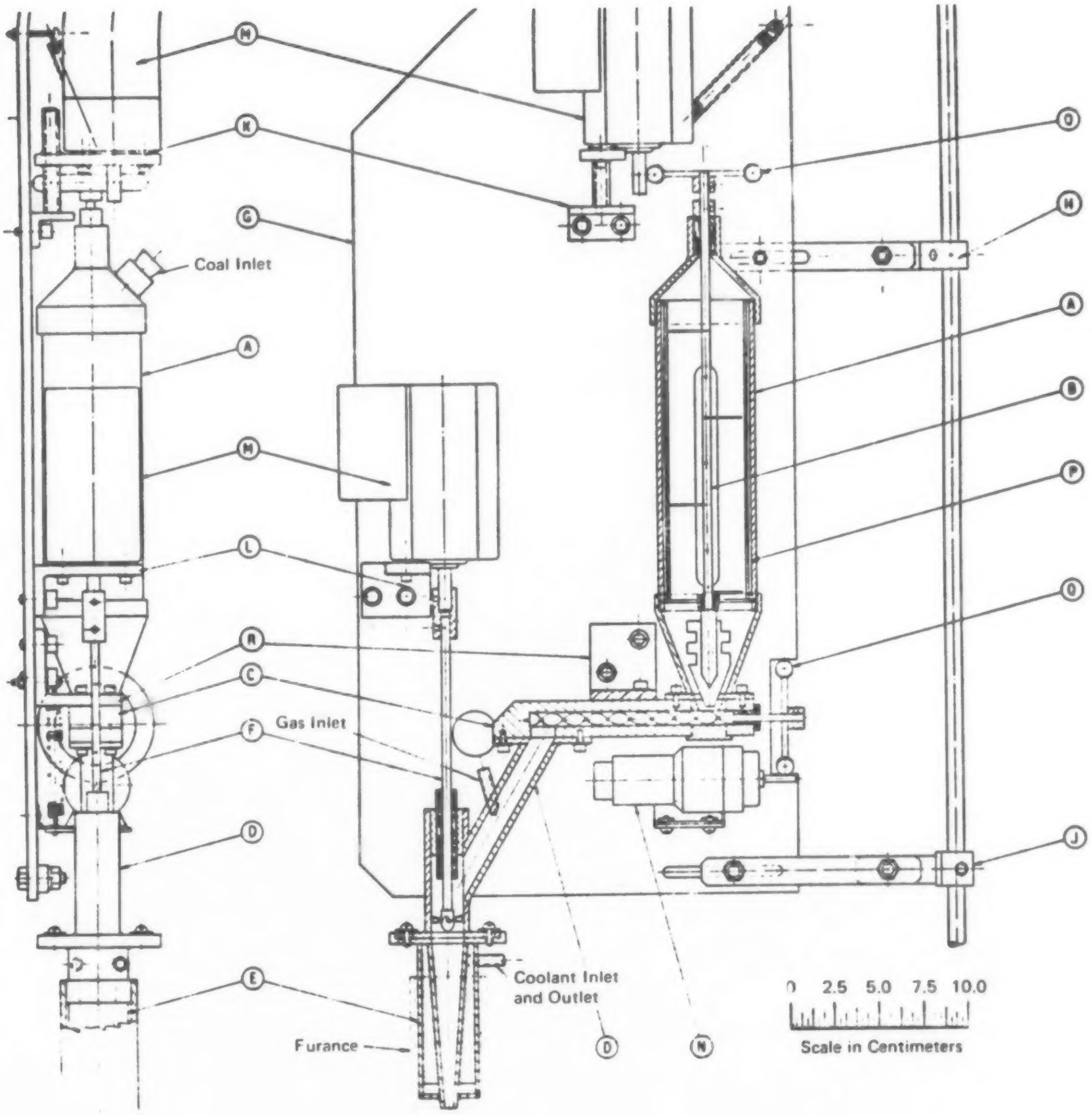


A	Hopper With Level
B	Stirrer
C	Screw Feed
D	"Y" Junction
E	Nozzle Feed
F	Mixer
G	Back Plate
H	Support Arm
J	Support Arm
K	Bracket
L	Bracket
M	Electric Motor
N	Electric Motor
P	Glass Tube Insert
O	Wheel
R	Bracket

Figure 1 The ACIRL micro-combustor.

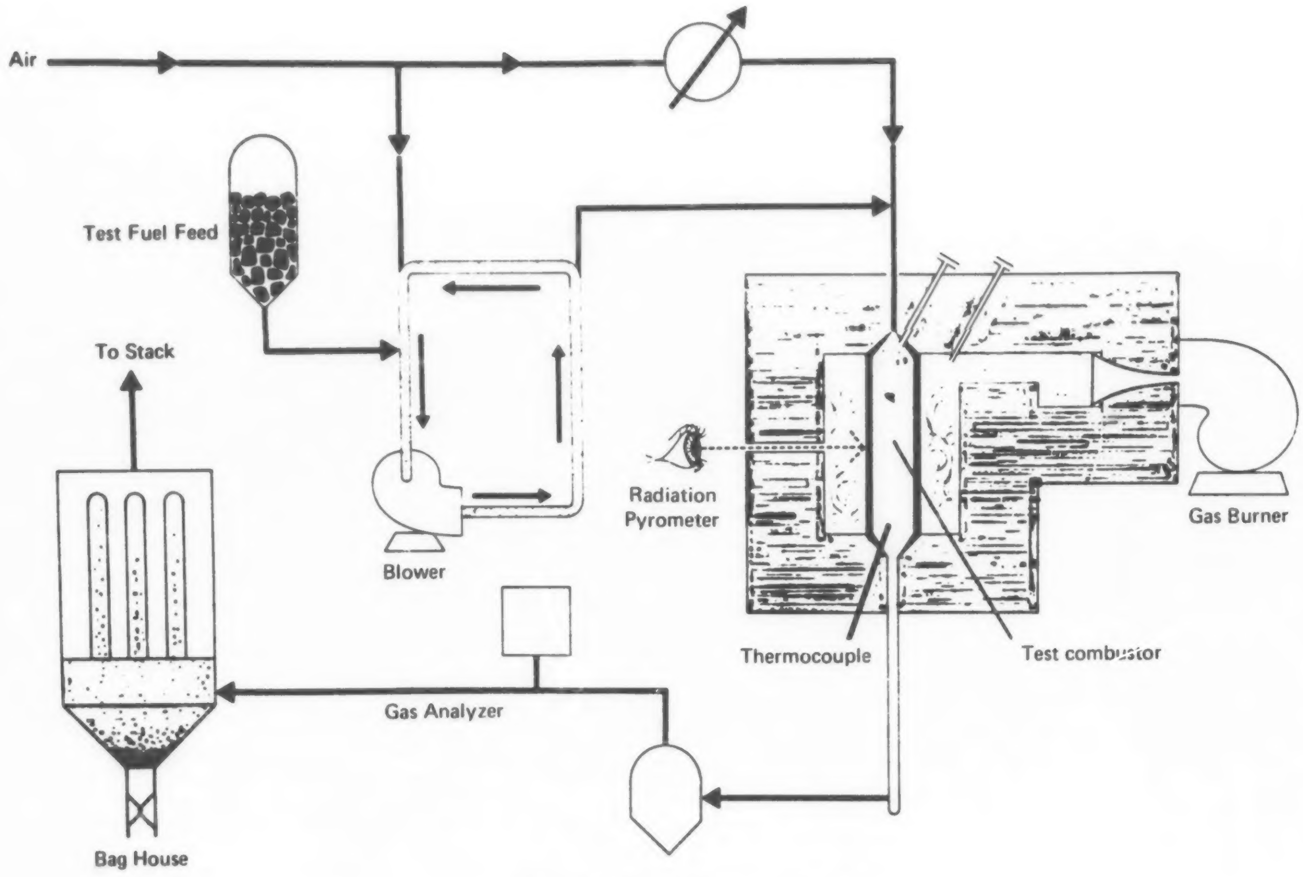


Figure 2 The Exxon combustor.

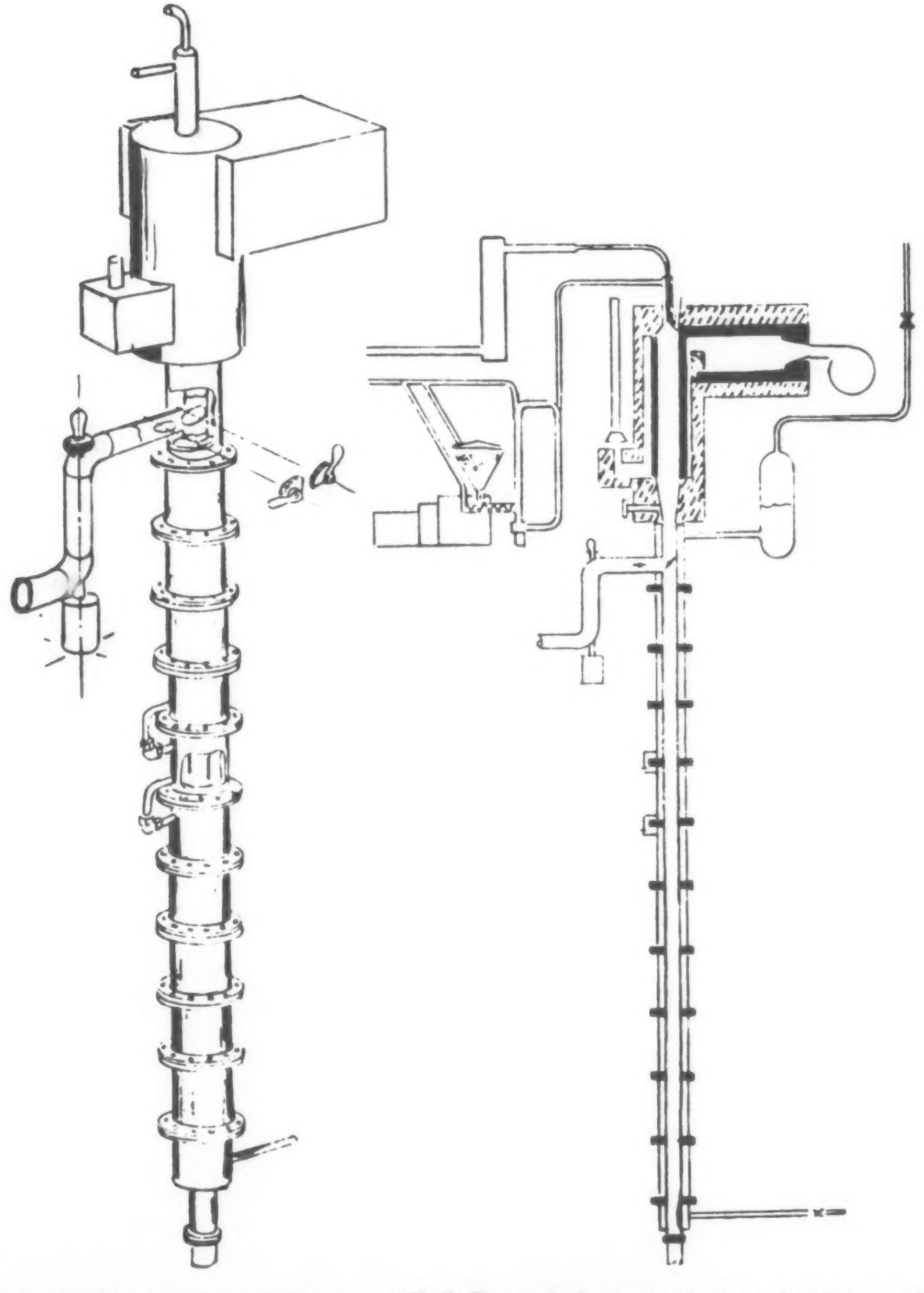


Figure 3 The B&W combustor (recently modified). The vertical segmented column is a water-cooled heat exchanger to cool the hot gases from the combustor. A test precipitator now planned will be installed in one of the segments where the gas temperature is at the desired level. B&W is also building a laboratory precipitator resembling the Australian "Kevatron".

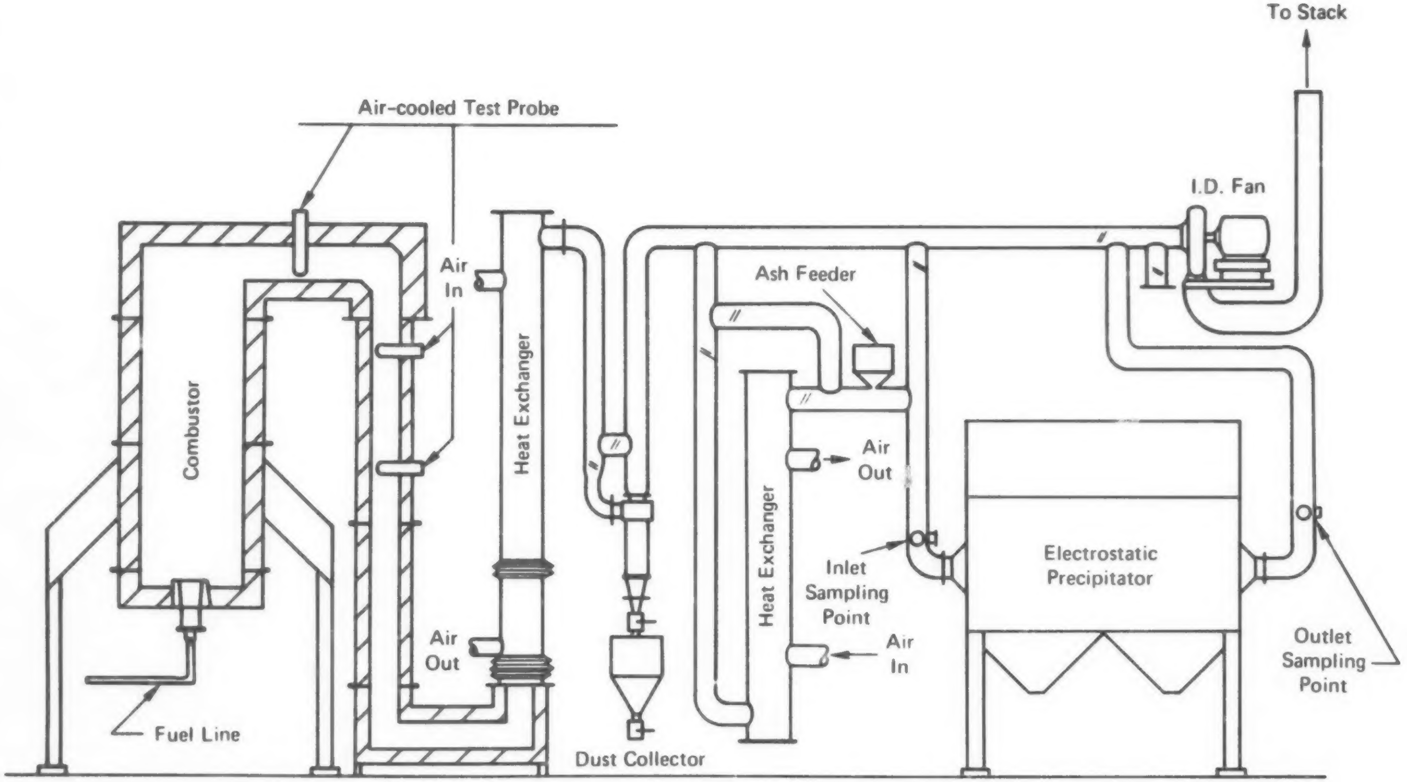


Figure 4 The Grand Forks combustor (rated 35 kg/hr coal) and electrostatic precipitator test system.

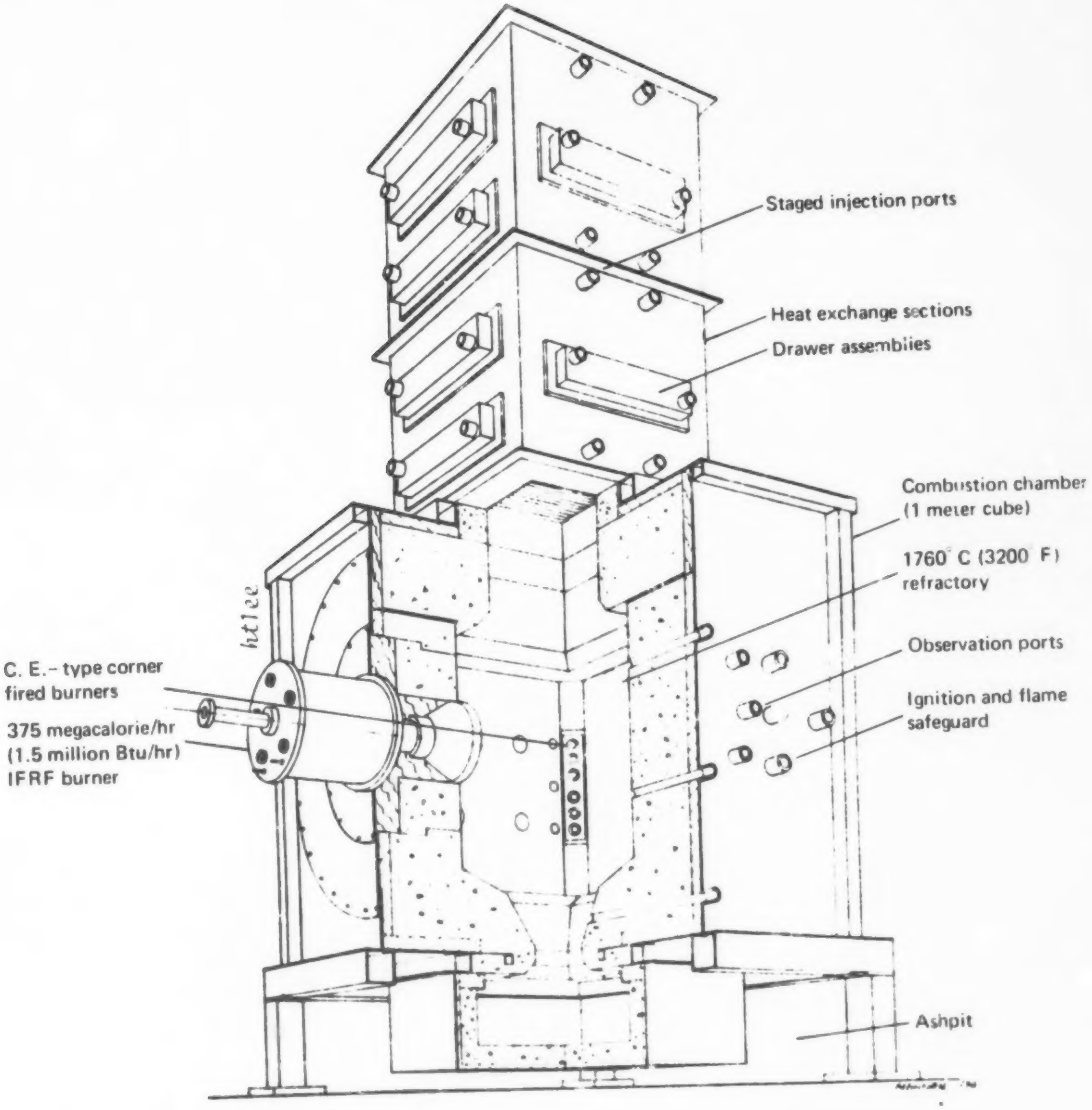


Figure 5 The Aerotherm-Acurex multi fuel combustor.

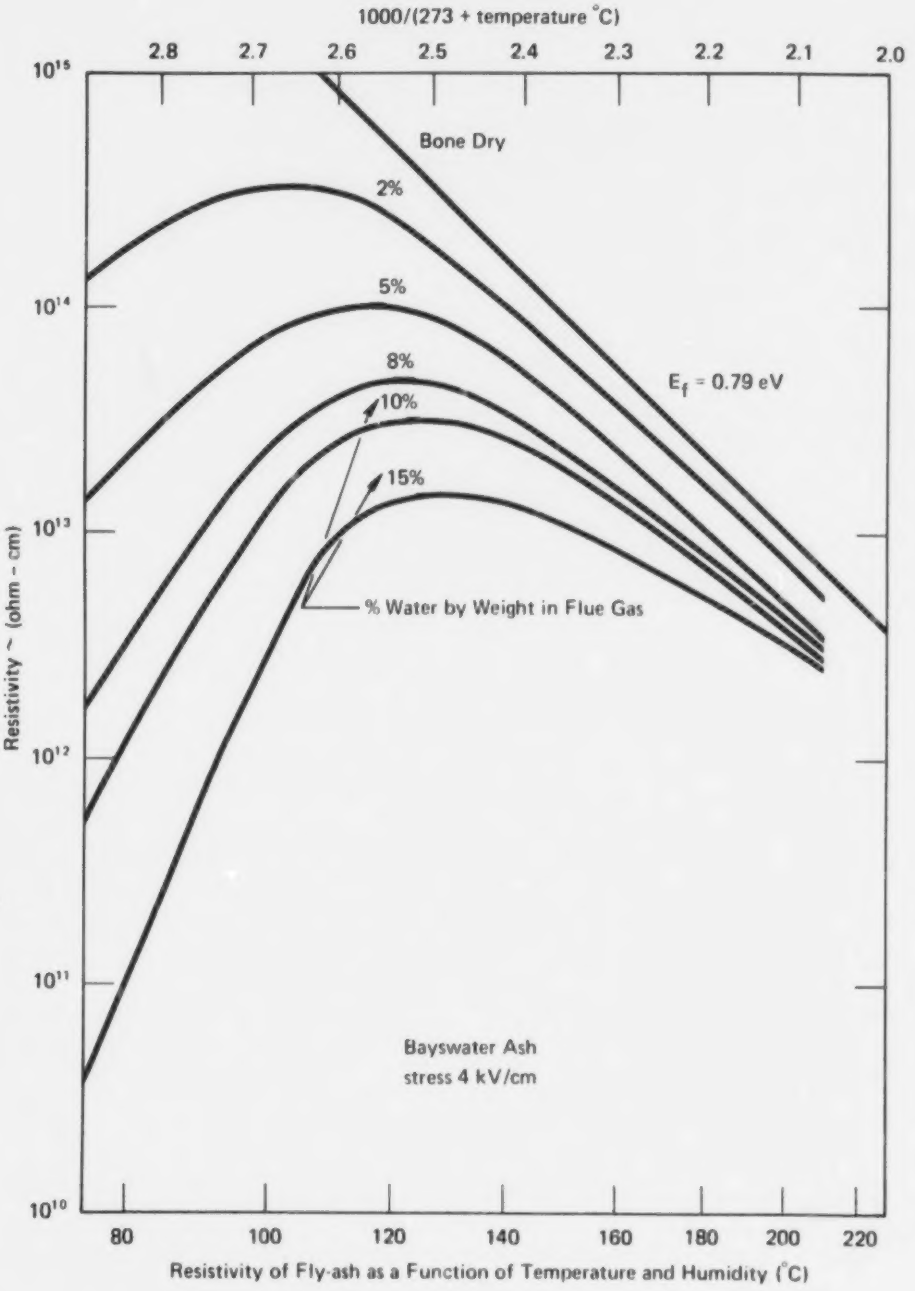


Figure 6 Typical resistivity curves for flyash from low-sulfur coal.

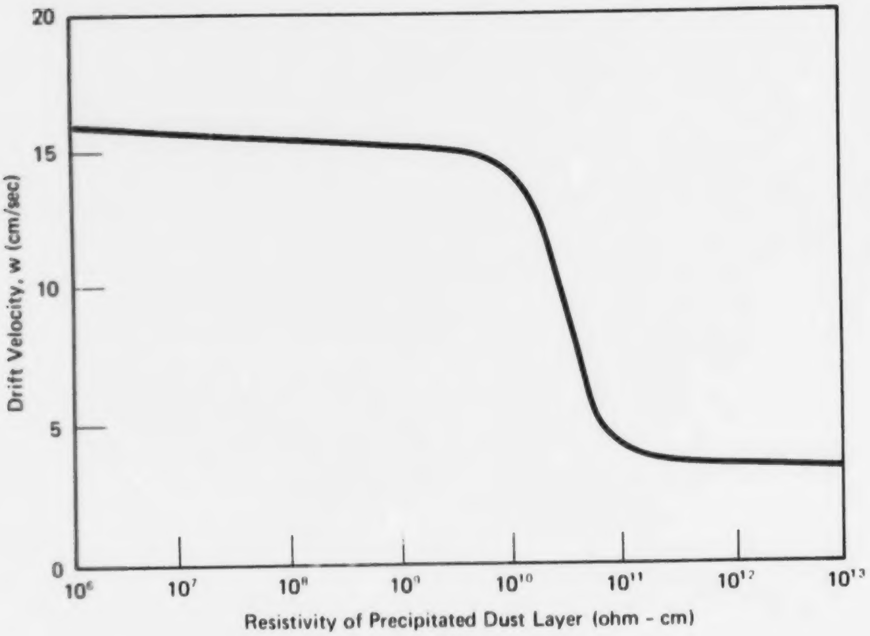


Figure 7 Curve showing general trend of conventional precipitator performance (w) as a function of dust resistivity.

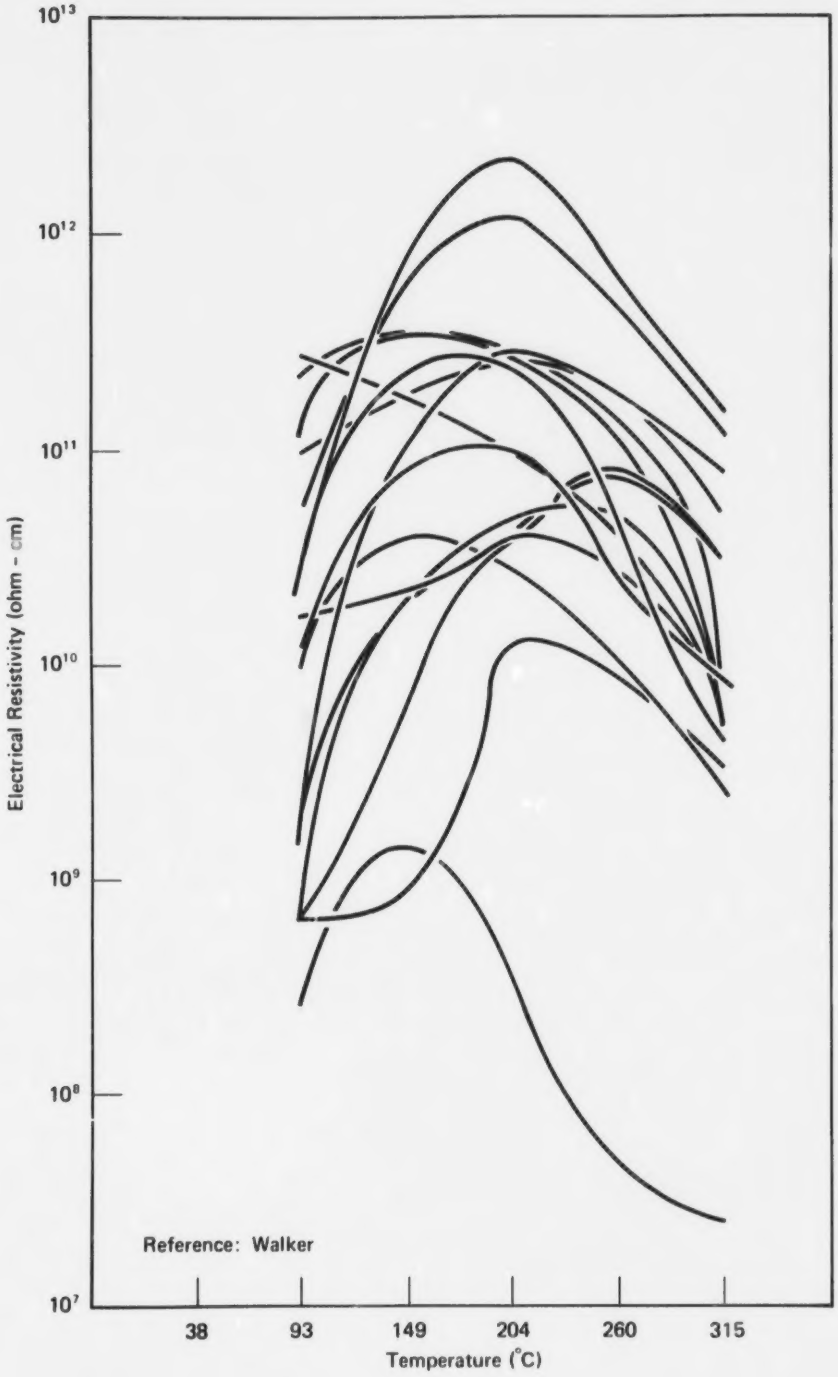


Figure 8 Laboratory resistivity curves for flyash samples from several Western coals. Stress 1 kv across 3 mm ash layer; 10 to 12% moisture by volume except 6% for one "hot" precipitator sample.

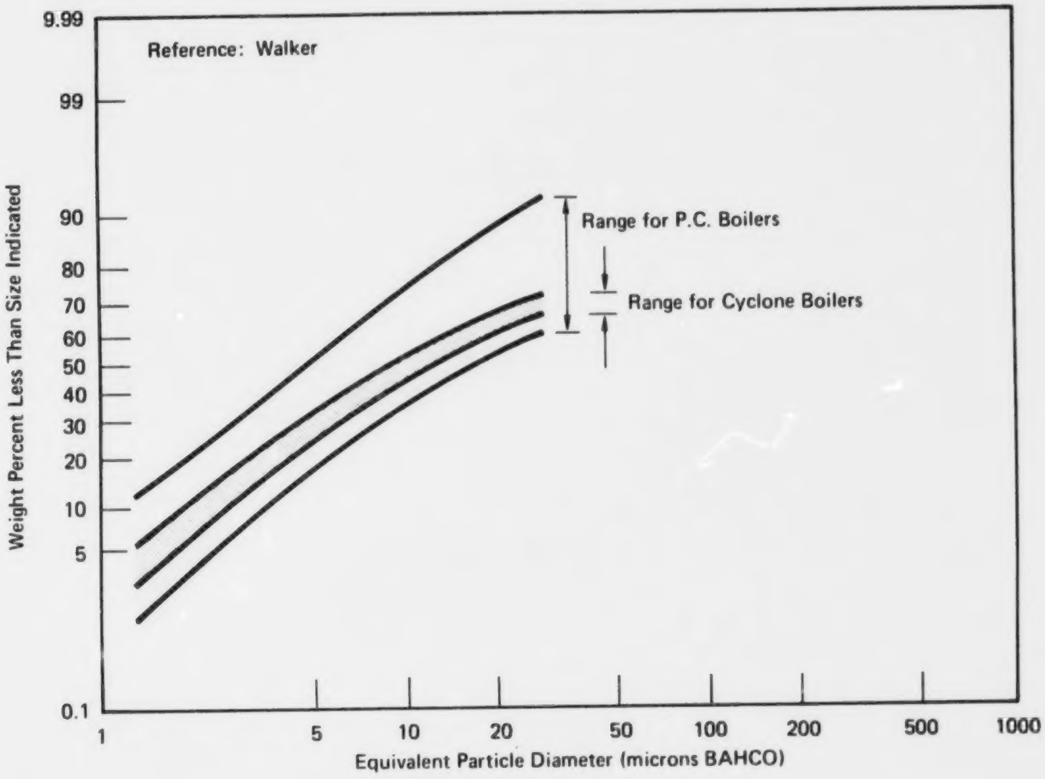


Figure 9 Ash particle size distribution curves approximate straight lines when plotted as shown here.

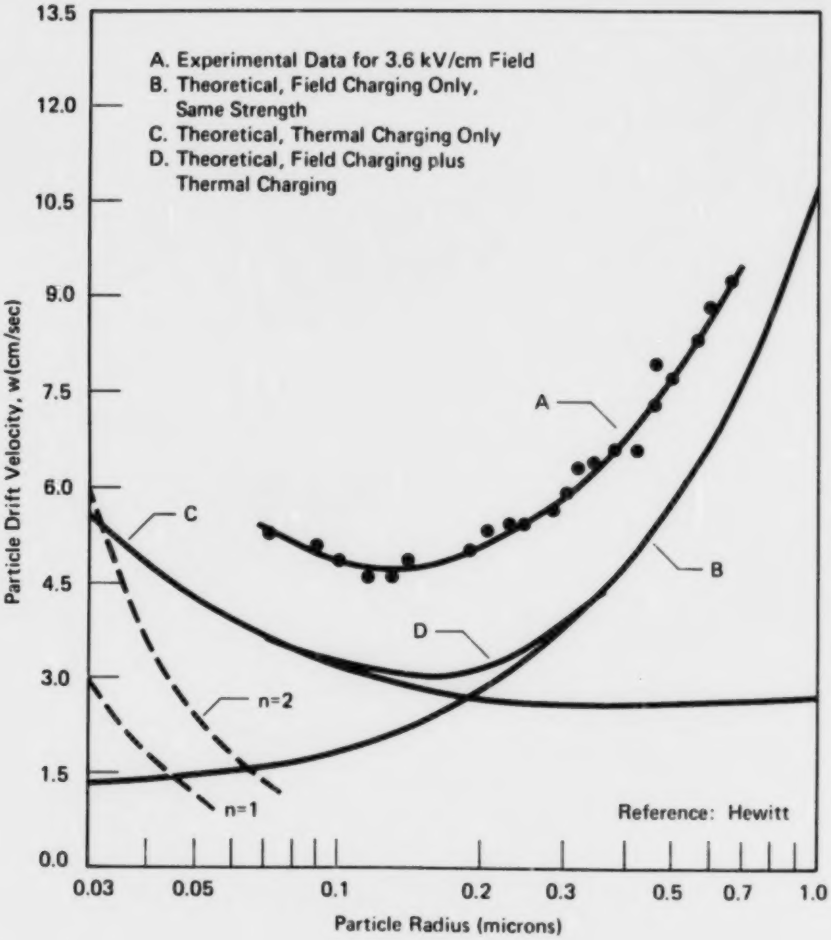


Figure 10 Drift velocity (w) as a function of particle size.

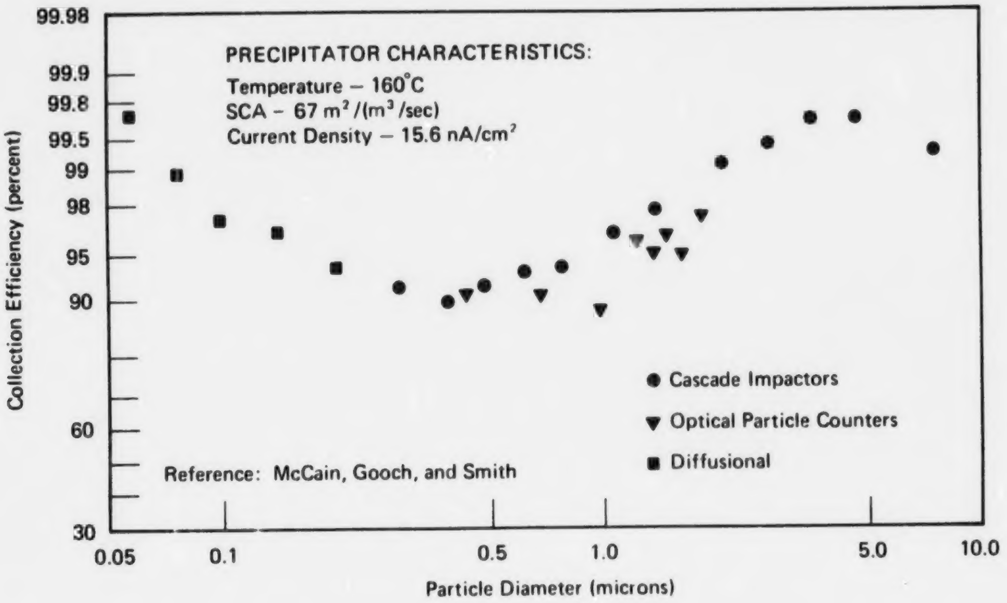


Figure 11 ("Cold") precipitator efficiency for varying flyash particle sizes.

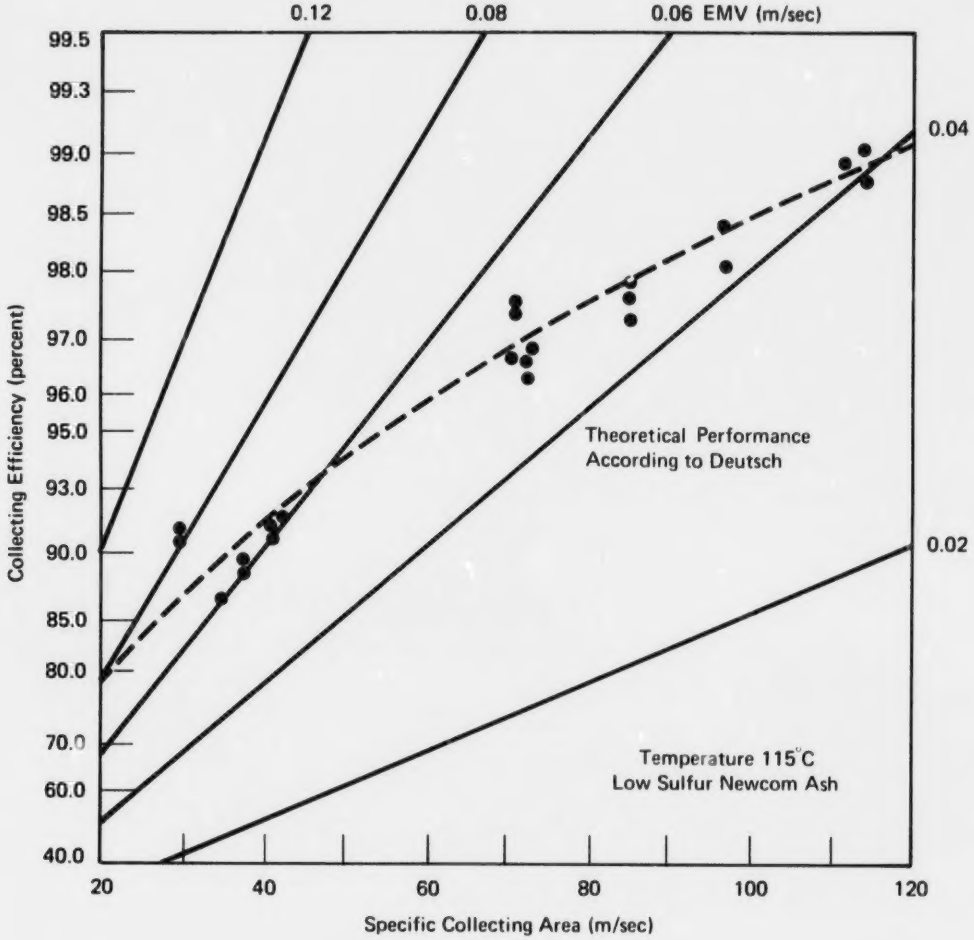


Figure 12 Collection efficiency as a function of f or SCA. Illustrating the fit of the regression equation to some experimental data.

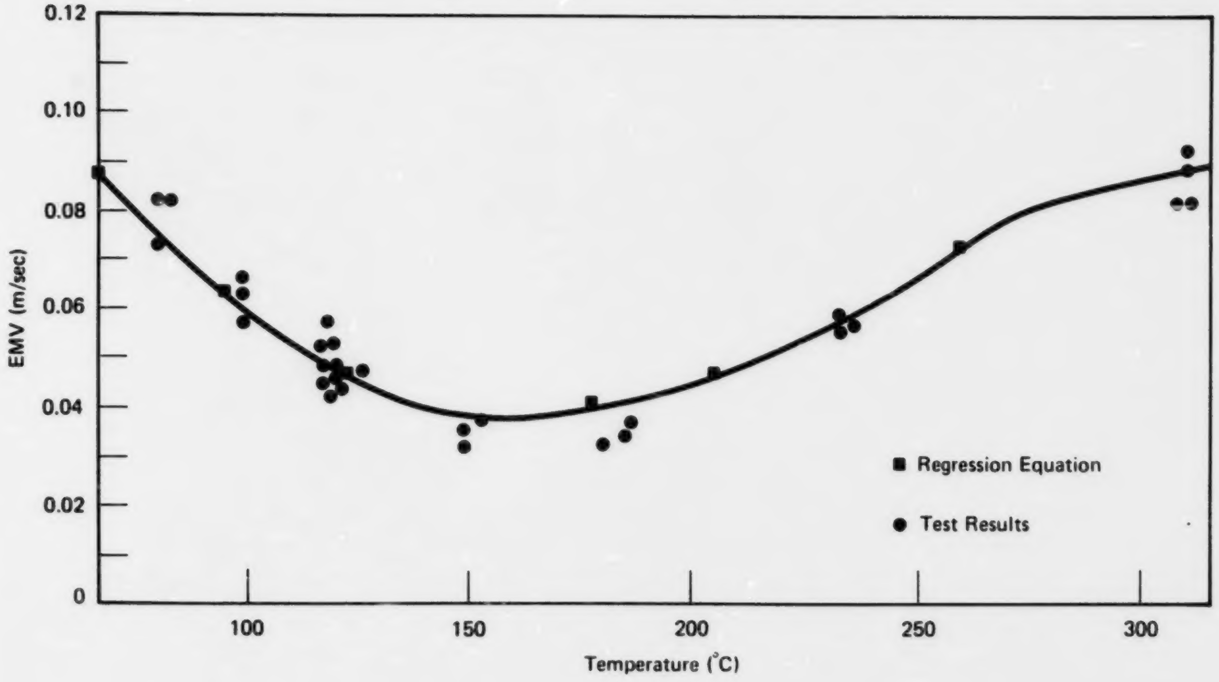
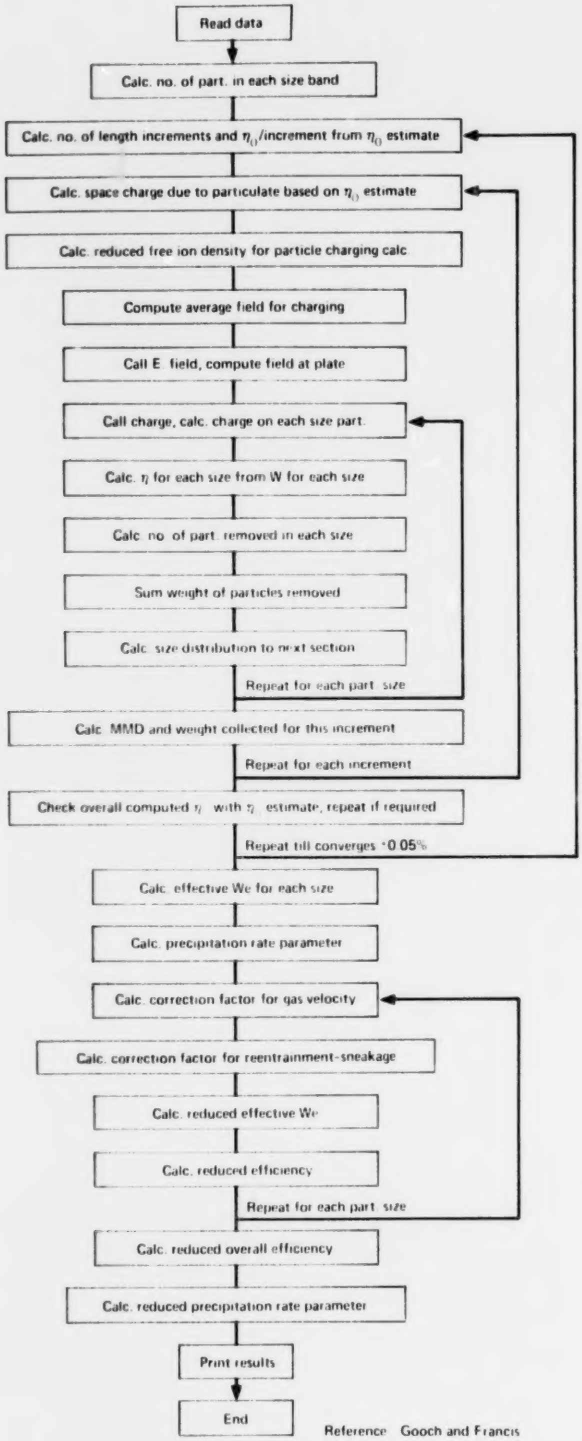


Figure 13 Precipitator performance as a function of temperature, illustrating the fit of the regression equation to the test data.


Read data
Calc. no. of part. in each size band
Calc. no. of length increments and η_0/increment from η_0 estimate
Calc. space charge due to particulate based on η_0 estimate
Calc. reduced free ion density for particle charging calc
Compute average field for charging
Call E field, compute field at plate
Call charge, calc. charge on each size part.
Calc. η for each size from W for each size
Calc no of part. removed in each size
Sum weight of particles removed
Calc size distribution to next section
Repeat for each part size
Calc MMD and weight collected for this increment
Repeat for each increment
Check overall computed η with η estimate, repeat if required
Repeat till converges ±0.05%
Calc. effective We for each size
Calc. precipitation rate parameter
Calc. correction factor for gas velocity
Calc. correction factor for reentrainment-sneakage
Calc. reduced effective We
Calc. reduced efficiency
Repeat for each part. size
Calc. reduced overall efficiency
Calc. reduced precipitation rate parameter
Print results
End

Reference Gooch and Francis

Figure 14 Simplified flow diagram of precipitator model computer program.

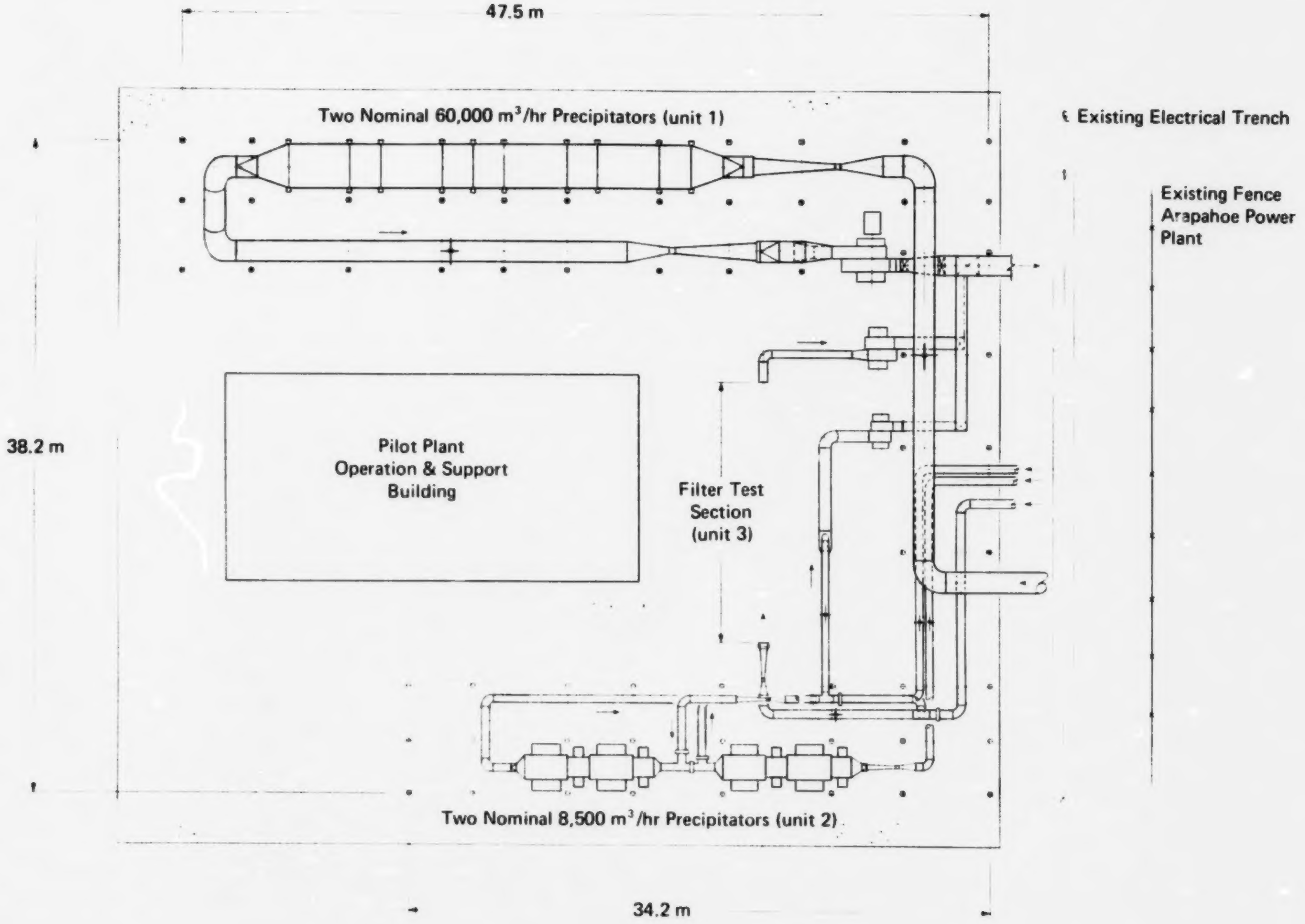

47.5 m
Two Nominal 60,000 m³/hr Precipitators (unit 1)
℄ Existing Electrical Trench
Existing Fence Arapahoe Power Plant
38.2 m
Pilot Plant Operation & Support Building
Filter Test Section (unit 3)
Two Nominal 8,500 m³/hr Precipitators (unit 2)
34.2 m

Figure 15 Preliminary plan of pilot precipitator experimental facility to be built at Arapahoe Station near Denver.

TECHNIQUES FOR CONDITIONING FLY ASH

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ABSTRACT

Conditioning of fly ash is discussed in connection with electrostatic precipitators from several points of view: the purposes served, the types of chemicals used, and their mechanisms of action. Major emphasis is given to resistivity modification with sulfur trioxide and other agents. Comments are also made on alternative conditioning mechanisms: increasing the cohesiveness of deposited fly ash to minimize reentrainment and improving the electrical properties of flue gas to increase the efficiency of collection.

TECHNIQUES FOR CONDITIONING FLY ASH

INTRODUCTION

Conditioning of fly ash usually involves the injection of a chemical (most often sulfur trioxide) into flue gas to control the electrical resistivity of the ash and improve its collection in an electrostatic precipitator. Currently, however, the term is assuming a much broader meaning, for conditioning involves processes other than resistivity modification. This paper discusses conditioning to achieve resistivity modification, but it also deals with two other purposes of conditioning: to increase the cohesiveness of fly-ash particles deposited in a precipitator and to improve the electrical properties of the gas stream flowing between discharge wires and collection electrodes.

RESISTIVITY MODIFICATION: TREATMENT OF ASH FROM LOW-SULFUR COAL

Conditioning by Sulfur Trioxide

Origin of High Resistivity. The production of fly ash with a high electrical resistivity is the usual—but not inevitable—consequence of the combustion of a low-sulfur coal. The problem of high electrical resistivity is clearly associated with the combustion of Western coals containing around 0.5% of sulfur. It is widely assumed to occur when Eastern coals of relatively low-sulfur content (often around 1%) are substituted for the more prevalent Eastern coals containing 3 to 4% sulfur. However, in terms of the frequently quoted upper limit of acceptable resistivity values (about 1×10^{10} ohm cm), ash from Eastern coals presents less of a difficulty than ash from Western coals. There are at least two factors that control resistivity: one is the sulfur content of the coal, and the other is the overall elemental composition of the fly ash.

Sulfur occurs in coal as organic compounds and as inorganic compounds—notably pyrite and sulfate salts. Regardless of the distribution of the sulfur in these different forms, however, the predominant product of combustion in the boiler is sulfur dioxide. From a thermodynamic point of view, sulfur dioxide would be the only product found in the boiler, and sulfur trioxide (or, more exactly, sulfuric acid) would be the only product present when the flue gas reaches a temperature around 150°C where electrostatic precipitators normally operate,[1] as shown in Table 1. Thermodynamics evidently is the process-limiting factor in the boiler but not toward the end of the flue-gas train, for the kinetics of oxidation does not allow more than a small fraction of the sulfur oxides to appear as sulfur trioxide.

Table 1 EQUILIBRIUM DISTRIBUTION OF OXIDIZED FORMS OF SULFUR AT VARIOUS GAS TEMPERATURES[a]

Temperature, °C	Relative concentrations, %, of sulfur compounds		
	SO_2	SO_3	H_2SO_4[b]
1400	99.7	0.3	0.0
1000	97.6	2.4	0.0
800	88.7	11.3	0.0
600	42.2	57.7	0.1
500	12.3	87.3	0.4
400	1.6	94.8	3.6
300	0.1	59.3	40.6
200	0.0	6.5	93.5
100	0.0	0.0	100.0

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a. Calculated from the data in JANAF Tables,[1] assuming concentrations of oxygen and water vapor equal to 4% and 10% by volume, respectively.

b. The maximum absolute concentration of each compound is sharply limited below 300°C as a result of the condensation of the predominant compound, H_2SO_4, in a binary H_2SO_4-H_2O liquid mixture.

Even so, the small fraction of the total of sulfur oxides occurring as sulfur trioxide at 150°C can be sufficient to lower the resistivity of ash to an acceptable range. Obvious factors of importance are the available concentration of sulfur dioxide (roughly proportional to the amount of sulfur in the coal) and the extent of oxidation to sulfur trioxide. A less obvious factor is the nature of the interaction of sulfur trioxide with the fly ash. We are only beginning to gain answers to this last question, but this much seems clear: sulfur trioxide and water vapor are jointly adsorbed or condensed on the surfaces of fly-ash particles, and either vapor assists in the deposition of the other. The resulting acid may remain essentially intact in a chemical sense, producing a conductive surface layer on an otherwise poorly conducting substrate. The acid may, however, react with basic constituents of the ash and thus undergo conversion to a nonconducting layer of sulfate salts. Calcium oxide, a fairly abundant component of ash from Western coals, is a probable cause of acid neutralization and its nullification as a conductor. Thus, two coals similar in sulfur content may ultimately produce similar concentrations of sulfur trioxide, but the one producing a more alkaline ash is likely to have a substantially higher resistivity.

Conditioning by Naturally Occurring Sulfur Trioxide. Some of the concepts outlined above were verified[2] in a study of the natural conditioning of fly ash in a plant burning an Eastern coal containing about 2% of sulfur and producing an ash in which the total percentage of alkaline oxides was low, about 7% by weight (4.2% as the alkali-metal oxides Li_2O, Na_2O, and K_2O, and 2.4% as the alkaline-earth oxides MgO and CaO). Experimental data showing fly-ash resistivity in a range of temperatures at the precipitator inlet are shown in Figure 1. The upper curve shows the expected maximum—in this instance, around 175°C—that occurs as the result of the transition between volume conduction at high temperatures and surface conduction at low temperatures. The lower segment of a different curve indicates the effect produced by injecting water vapor in an amount sufficient to double the normal level—i.e., increasing the concentration of water vapor from 7% to 14% by volume.

Other experimental data of value in explaining the observed changes in resistivity are given in Figures 2 and 3. Measured concentrations of sulfur trioxide in gas samples taken upstream and downstream from the air preheater indicate the occurrence of two effects: (1) the thermodynamically predicted increase in the oxidation of sulfur dioxide to sulfur trioxide as the gas temperature falls across the air heater (but short of the extent predicted for equilibrium) and (2) the loss of sulfur trioxide from the gas phase, a phenomenon aided by either the lowering of temperature or the injection of water vapor. Measurements of slurries of fly ash in water show corresponding increases in the acidity of the ash (decreases in pH) and in the sulfate content of the ash.

Conditioning by Injected Sulfuric Trioxide. In plants where the naturally available concentration of sulfur trioxide is too low—in an absolute sense, as the result of a low sulfur concentration in the coal or, in a relative sense, as the result of reaction with a highly alkaline ash—the conditioning agent can be introduced by an artificial process. Sources of sulfur trioxide are vapor generated by vaporizing the liquid form of this compound, vaporizing sulfuric acid, or oxidizing sulfur dioxide gas in a vanadium pentoxide catalyst. In the last-mentioned type of process, sulfur dioxide may be purchased as the liquid compound under pressure and evaporated prior to conversion, or it may be generated at the site in a burner for elemental sulfur. The molecular composition of vapors injected in the various systems depends not only on the source material but on the temperature and water vapor concentration in the dilution air. The vapor may be sulfur trioxide per se, or it may be sulfuric acid. It is of little consequence which substance is injected for, once injected, either compound will occur only as sulfuric acid vapor in the gas stream. This statement stems from the rapid rate of the gas phase reaction $SO_3 + H_2O \longrightarrow H_2SO_4$ under typical flue-gas conditions (temperatures around 150°C and water-vapor concentrations around 7% to 10% by volume), to approach equilibrium conditions as indicated in Table 1.

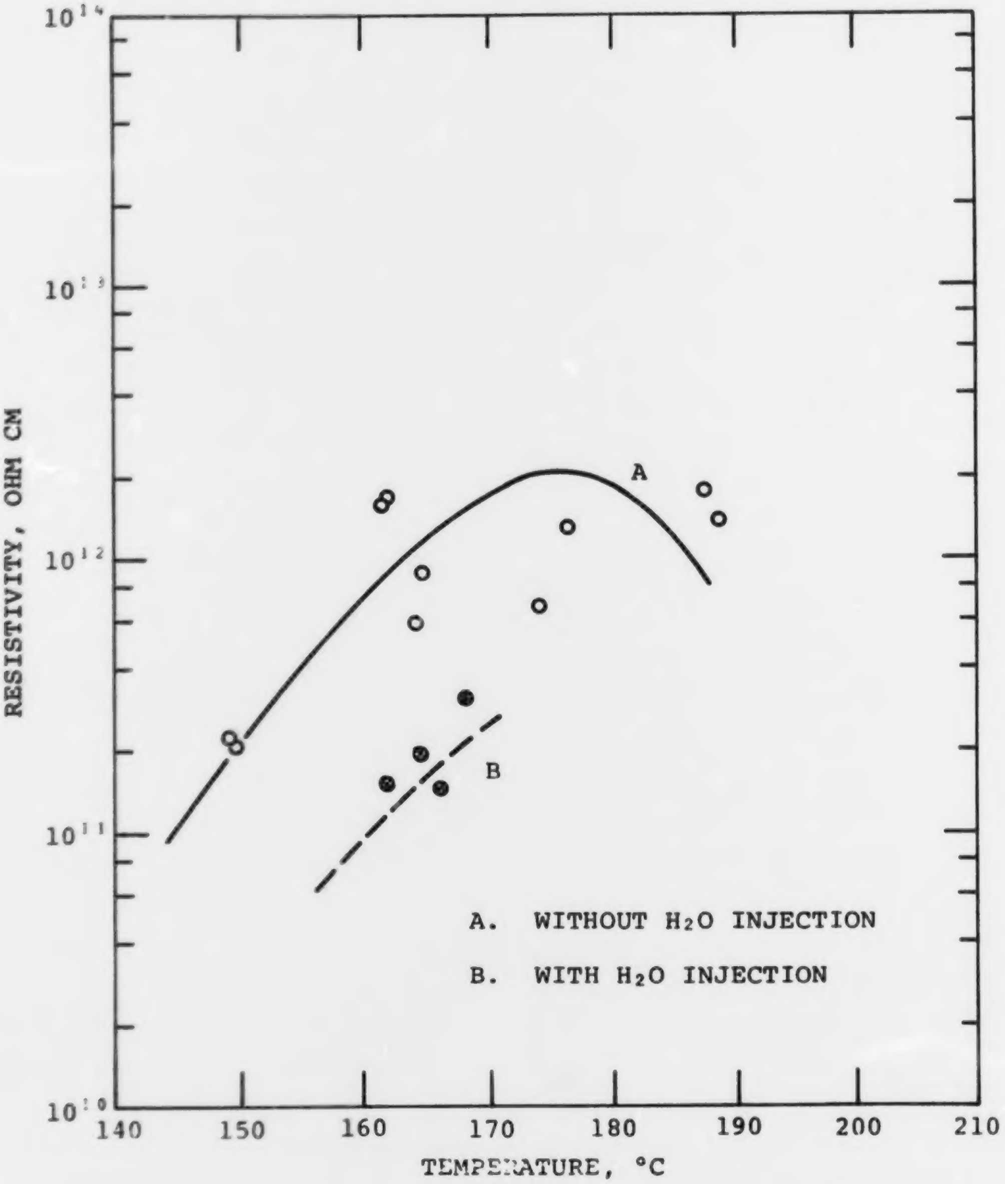

Figure 1 Electrical resistivity of fly ash as a function of temperature or water vapor concentration.

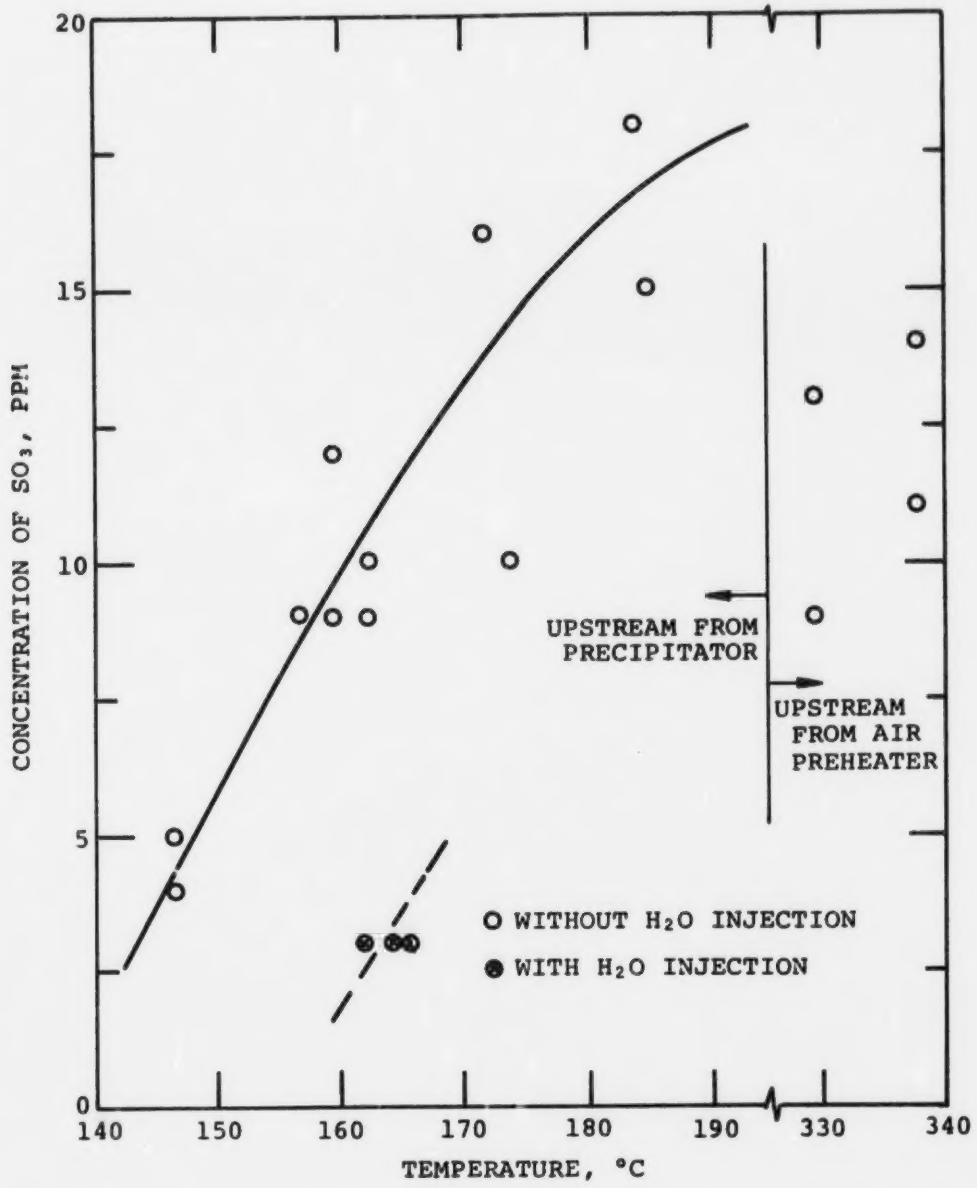

Figure 2 Concentration of sulfur trioxide as a function of sampling location or temperature.

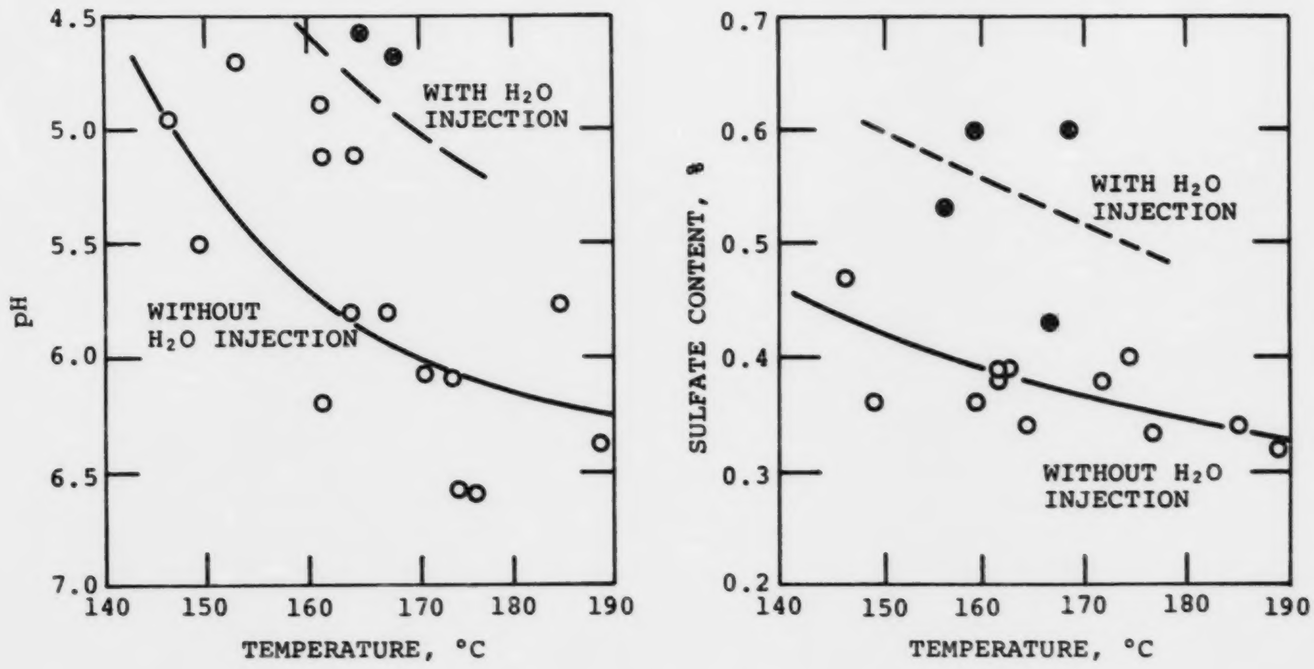

Figure 3 Acidity and sulfate content of fly ash as functions of temperature and water vapor concentration.

Investigators at the Institute have studied the results achieved by the injection of sulfuric trioxide or sulfuric acid in a number of full-scale power plants.[2] We have considered both practical and theoretical questions and obtained answers that, to us, seem generally satisfactory, subject to the limitations imposed by the characteristics of plants available for study.

One question considered is whether there is any high-resistivity ash that cannot be suitably conditioned with sulfur trioxide. We believe that the answer is no, but we do recognize that the quantity of conditioning agent required varies with such factors as the chemical composition of the ash and the gas temperature. The importance of these factors is illustrated in Figure 4, which gives the results obtained with ashes characterized broadly as acidic, neutral, or alkaline at temperatures varying from 110 to 160°C. If a resistivity of about 1×10^{10} ohm cm is required, more sulfur trioxide is required if the ash is alkaline or if the gas temperature is high.

Another question considered is whether the efficiency of conditioning is markedly affected by the choice of location in the flue-gas train where the agent is added. Our studies have included plants with and without mechanical collectors to aid precipitators in the collection of fly ash; they have included plants with injection ahead of the mechanical collectors, between the mechanical and electrostatic collectors, and ahead of precipitators operating without mechanical collectors. No important change in the efficiency of conditioning can be attributed to these variations in plant operation. We have not had experience with injection ahead of the air preheater, which would be attractive from the point of view of ensuring uniform introduction of sulfur trioxide or sulfuric acid as a vapor.

Another question was whether a choice is to be made among the different types of injection systems (based on injection of sulfur trioxide directly, as catalytically oxidized sulfur dioxide, or as sulfuric acid). We have not recognized any change in the efficiency of conditioning that we believe to be fundamentally attributable to the type of system. We did have experience with one acid-injection system that was inefficient as the apparent result of excessive condensation near the point of injection.

A fourth matter of practical interest was whether a significant fraction of the injected sulfur trioxide is lost through the stack. A close examination of this point was made in only one plant where conditions favored a stack loss, as the result of the acidic character of the ash and incomplete removal of the conditioning agent. Figure 5 compares the measured concentrations of sulfur trioxide with and without injection as a function of distance across the outlet duct at the precipitator or, in

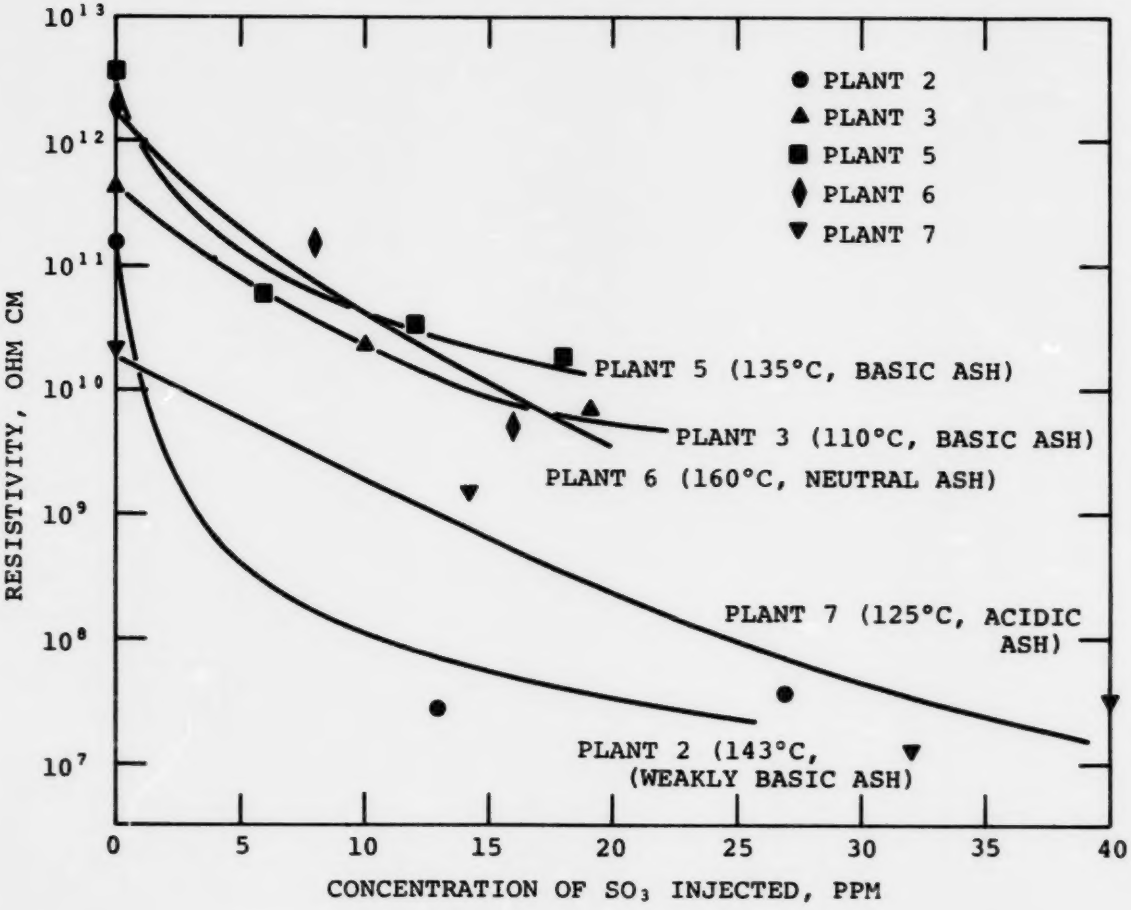

Figure 4 Resistivity as a function of the concentration of injected sulfur trioxide.

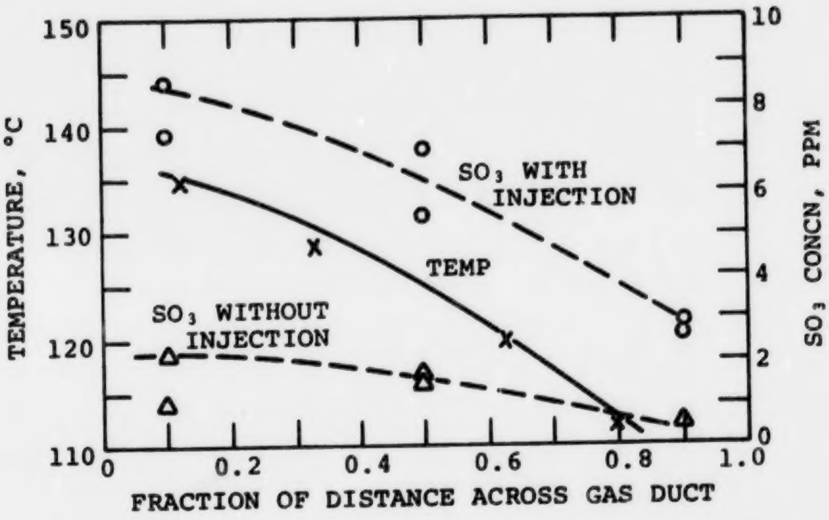

Figure 5 Concentration of sulfur trioxide as a function of gas temperature at the outlet of a precipitator.

other words, as a function of the gas temperature (a variable across the duct as a result of the influence of the Ljungstrom air heater). The difference in the areas under the two concentration curves corresponds to an average concentration of 5 ppm, about 30% of the 14 ppm injected. An increase in the sulfur trioxide concentration by 5 ppm would produce a concentration of condensed sulfuric acid of 20 μg/m³ (about 0.01 gr/ft³) in the relatively cool plume from the stack—a significant level in terms of "clear stack" concentration levels for fly ash. We recognize that, under typical conditions with an alkaline ash from low-sulfur Western coals, the stack loss of sulfur trioxide will be lower and perhaps insignificant. But it is not a matter that can be ignored.

A theoretical question of some importance is: What is the mechanism of conditioning? One facet of this question is whether deposition of the agent into ash occurs by adsorption or condensation. Our conclusion is that adsorption of vapor above the acid dew point can certainly occur, although condensation may also occur if the gas is below the dew point. This conclusion is based on a prediction of dew points from thermodynamic data that are applicable only if the condensed vapor occurs as a binary liquid mixture of sulfuric acid and water. The possibility that reaction of vapors with alkaline fly-ash constituents may take precedence over condensation makes prediction of dew points on this basis uncertain; if such a reaction occurs, however, the deposition of vapors would be even more likely to occur as an adsorption process.

Conditioning by Other Agents

No discussion of conditioning of high-resistivity fly ash would be complete without mention of the role of ammonia, although the subject must be approached with caution. It appears that this agent sometimes is effective by other mechanisms of conditioning, as discussed subsequently, and that its ability to alter resistivity is not clear-cut.

The value of ammonia as a conditioning ash for one type of suspended particulate matter—catalyst dust in a petroleum refinery, rather than fly ash from a coal-burning boiler—has been known for many years. J. F. Chittum of the former Western Precipitation Company was one of the pioneers in the use of ammonia. He believed that ammonia was of greater value than sulfur trioxide in treating an acidic dust, such as the catalyst material in a refinery.[3]

We may very well ask: What can be expected of ammonia as a conditioning agent for fly ash? Suppose we accept Chittum's premise that the use of ammonia and not sulfur trioxide is indicated if the ash is acidic. We would then have to imagine a fly ash that is at once acidic and high in resistivity. The experience and interpretation of ash chemistry leads to the viewpoint that these may be mutually exclusive factors. None of the constituents of fly ash except adsorbed sulfur trioxide is sufficiently acidic to have a strong affinity for ammonia, which is a rather weakly basic compound. But, if adsorbed sulfur trioxide is present, how can resistivity be high?

These are ideas that will have to change if enough clear-cut evidence of a resistivity effect by ammonia becomes available. There are data available either directly indicating that ammonia lowers resistivity or indirectly indicating that ammonia has this effect, such as changes in precipitator voltages and currents that are most easily explained as a result of lowered resistivity. It does not appear, however, that conclusive evidence of lowered resistivity with ammonia conditioning is now in existence.

What about other agents? Sodium is an example that we believe to have clear-cut significance in lowering resistivity. The importance of sodium first became evident in a comparison of resistivity values of ashes varying naturally in percentages of sodium oxide. It was later confirmed in trials of various sodium compounds (notably the carbonate) as additives to coal being fed to the boiler of a power plant. Our experience on the use of sodium is entirely restricted to boiler additions (not flue-gas treatment),[4] but others have reported the use of sodium as a flue-gas additive.[5]

What about some of the proprietary agents now on the market? For some of these agents, the claim is made that they function as resistivity modifiers, but the data upon which such claims are made are not often, if ever, given. From what is known about the compounds present in the proprietary agents and what has been published from laboratory studies of the compounds, their functioning as resistivity modifiers is not an unreasonable assumption.[6] One of the compounds identified is sulfamic acid; another is ammonium sulfate. It is perhaps noteworthy that each of these compounds is, in theory at least, derived from sulfur trioxide and ammonia. From a constitutional point of view, sulfamic acid is equivalent to these constituents in a 1:1 mole ratio ($HO\text{-}SO_2\text{-}NH_2 = SO_3 + NH_3$). Similarly, ammonium sulfate or $(NH_4)_2SO_4$ is equivalent to sulfur trioxide, ammonia, and water in the molar proportions $SO_3:NH_3:H_2O = 1:2:1$. Thermal decomposition of either compound can thus produce both sulfur trioxide and ammonia as gases along with further breakdown products, such as SO_2 and N_2.

RESISTIVITY MODIFICATION OF ANOTHER TYPE: TREATMENT OF ASH FROM HIGH-SULFUR COAL

It is important to realize that the problem of excessively low resistivity occasionally exists, although it is less widely recognized than the problem of high resistivity. It occurs in the burning of a high-sulfur coal, and it conceivably may occur if ash from a low-sulfur coal is overtreated with sulfur trioxide. If resistivity is low—below 1×10^7 ohm cm—the practical consequence may be excessive reentrainment, owing to the inadequacy of the electrical force across the ash deposit for maintaining the physical integrity of the deposit under the influence of scouring or electrode rapping.

In view of the likelihood that low resistivity can be attributed to an excessive amount of sulfur trioxide, it is reasonable to try ammonia as an agent for chemically neutralizing the acid. From a theoretical point of view, one must be troubled over the prospects of increasing resistivity sufficiently by simply converting sulfuric acid to ammonium sulfate. The conductivity of ammonium sulfate in water is lower than that of sulfuric acid, but the ratio of conductivities is roughly 1:4 and thus not as great as one order of magnitude. A study of the mechanism of ammonia conditioning under circumstances where it apparently gave the desired resistivity change failed to confirm the mechanism assumed.

REENTRAINMENT CONTROL: INCREASING THE COHESIVENESS OF PRECIPITATED ASH

If, as just discussed, the resistivity of ash from a high-sulfur coal can be increased, one has a means for minimizing reentrainment. If one is unable to make any significant change in the resistivity and thus in the electrical force needed to maintain physical integrity of deposited ash, he has the option of increasing mechanical forces that bind individual particles together.

Dalmon and Tidy[7] recognized the value of sulfur trioxide as an agent for increasing the binding forces between particles of fly ash and unburned carbon. Carbon is a highly conductive material (relatively speaking) and is especially difficult to retain after it is electrostatically precipitated, except by means of cohesive forces. The method used by Dalmon and Tidy for demonstrating the reported activity of sulfur trioxide involved packing a bed of ash and carbon in a bed with one end detachable from the other and determining the mechanical force needed to rupture the bed of particles.

Investigators at the Institute[2] obtained evidence in one plant for a similar effect of sulfur trioxide, although in this instance there was no indication of excessive carbon being the cause of excessive reentrainment. In this plant, conditioning by sulfur trioxide had been adopted as a measure for restoring the original efficiency of a precipitator after a change had been made to burn coal with a lower sulfur content. Our data failed to indicate that resistivity was a problem; the resistivity of untreated ash was found to be about 1×10^{10} ohm cm, and the value with 14 ppm of sulfur trioxide injected was about 1×10^{9} ohm cm. One might have expected increased reentrainment as a result of the change in resistivity. In actuality, however, we found that injection of sulfur trioxide increased the precipitator efficiency markedly. Data in Table 2 show a parallel between the effects achieved by injecting sulfur trioxide or discontinuing electrode rapping. Our conclusion, therefore, is that the principal mechanism of conditioning in the circumstances investigated was increased cohesiveness of the fly ash and decreased rapping reentrainment.

Table 2 REDUCTION OF RAPPING REENTRAINMENT
BY SULFUR TRIOXIDE

SO_3 Injected, ppm	Rapping	Precipitator efficiency, %
0	Normal	69.7
	Off	82.0
14	Normal	93.5
	Off	96.8

We also found evidence that ammonia injection suppressed rapping losses in plants burning high-sulfur coal.[2] A reproduction of a chart recording the signal from a transmissometer at the precipitator outlet in one of these plants is given in Figure 6. After ammonia injection was started, the intensity of rapping puffs was slowly suppressed. After ammonia injection was discontinued, the intensity of rapping puffs gradually increased again. The findings at the plant in question would not be complete without mention of other experimental results: no measurable effect of ammonia on the resistivity of the ash (about 1×10^9 ohm cm) was evident, but significant effects of ammonia on the electrical properties of the gas stream were discerned, as discussed later.

It appears that triethylamine, which has recently been found a promising agent in Australia,[8] may act in the manner described for ammonia but even more effectively. Both agents are basic, but the basic strength of triethylamine is greater and the reported stronger effect of triethylamine seems logical. It must be recognized that the Australian workers have concluded that triethylamine acts as an "agglomerating" agent, which we interpret to mean an agent that increases particle size before precipitation occurs. However, if this mechanism of conditioning does occur, specific evidence for the process does not appear to have been obtained.

ALTERATION OF ELECTRICAL PROPERTIES OF THE GAS STREAM: THE SPACE-CHARGE EFFECT

Conditioning through this mechanism has been observed in several power plants where ammonia was used for conditioning fly ash.[2,9] Two of the plants investigated burned high-sulfur coals and produced fly ash with resistivities around 1×10^8 to 1×10^9 ohm cm. Another two plants burned so-called low-sulfur coals from Eastern mines, but in these instances the sulfur level was around 1.0% and measured resistivities were about 1×10^{10} to 1×10^{11} ohm cm. Moreover, significant concentrations of sulfur trioxide were found in the flue gas—around 2 to 5 ppm, in contrast to usually undetectable levels (less than 1 ppm) in plants burning 0.5%-sulfur Western coals.

No measurable effect of ammonia on resistivity could be detected. However, the gradual change in the intensity of rapping puffs, indicating an increase in the cohesiveness of deposited ash, was detected clearly in one plant. Moreover, there were other changes that were virtually instantaneous. One effect was a change in the precipitator electrical conditions in the direction consistent with a decrease in the mobility of charge carriers in the gas phase, as shown in Figure 7. Another phenomenon that occurred rapidly was a marked enhancement in the concentration of suspended

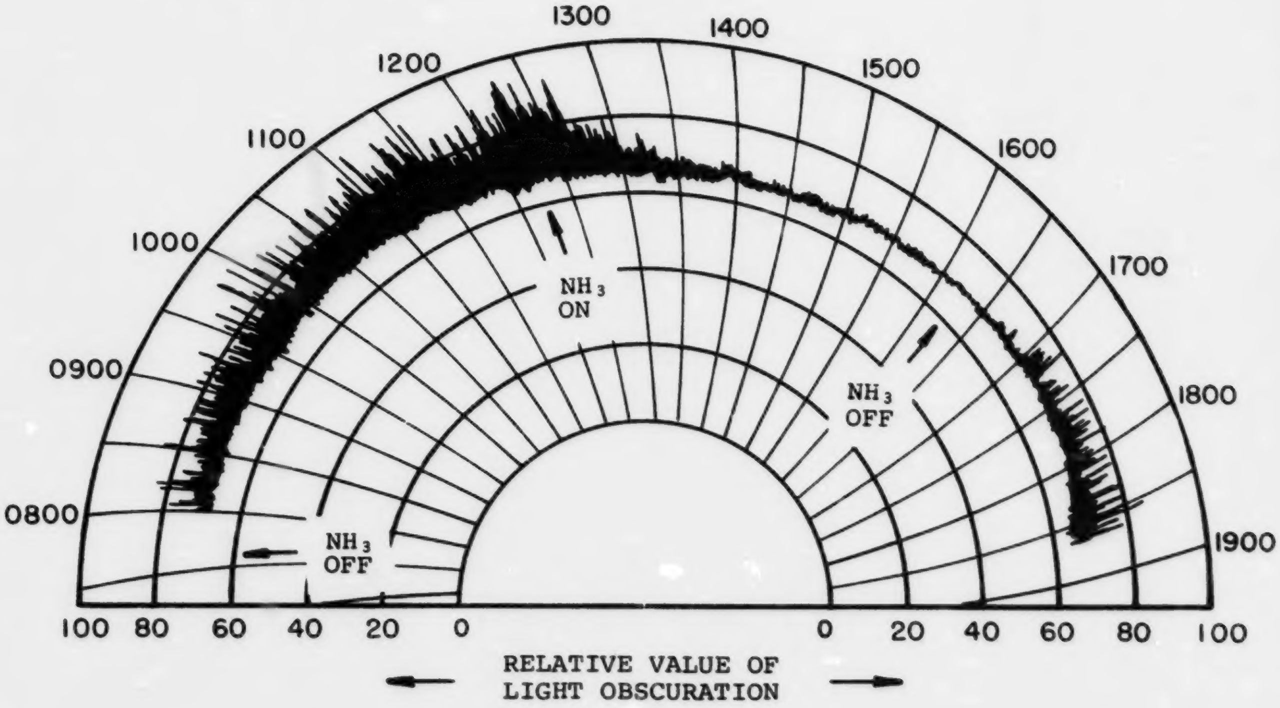

Figure 6 Reduction of rapping reentrainment by ammonia.

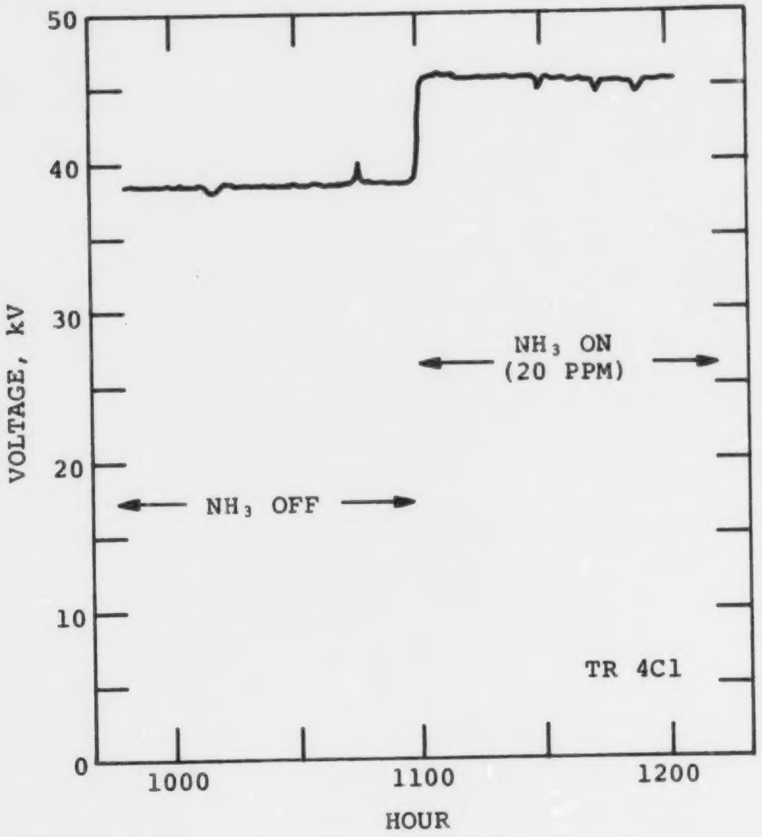

Figure 7 Rapidity of the effect of ammonia on the voltage of a precipitator.

fine particles at the precipitator inlet. Coupled with these effects were pronounced decreases in the concentration of sulfur trioxide as a constituent of the gas stream.

Taken together, the observations suggested a chemical reaction between the injected ammonia and the normally present sulfur trioxide (actually, the vapor of sulfuric acid) to produce a fume of fine particles of ammonium sulfate or perhaps the bisulfate, as shown by the following equations:

$$2NH_3(\text{gas}) + H_2SO_4(\text{gas}) \longrightarrow (NH_4)_2SO_4(\text{solid})$$

$$NH_3(\text{gas}) + H_2SO_4(\text{gas}) \longrightarrow NH_4HSO_4(\text{solid or liquid})$$

The bisulfate in the second equation is shown as either solid or liquid; the transition point is about 144°C.

What is the likelihood of a similar space-charge effect in a plant burning a typical low-sulfur Western coal, and what would be the value of such an effect? The reaction of injected ammonia with normally occurring sulfur trioxide would generally be less likely to occur in such a plant because of the insignificant quantity of sulfur trioxide present. However, much higher concentrations of ammonia than are generally used for conditioning—say, 100 to 200 ppm—could react with trace concentrations of sulfur trioxide (less than 1 ppm). Concentrations of ammonia in this range, it may be noted, are used to advantage with high-resistivity Australian ashes. Such concentrations of ammonia may also react with sulfur dioxide to produce fine particles of ammonium sulfite, although this does not appear likely from the thermodynamic data now available.

What about the possible value of the space-charge effect if it does occur? Possibly this effect could suppress the abnormal currents that are encountered in instances of back corona. Data published by Watson and Blecher in Australia some years ago[10] gave evidence of back corona suppression by ammonia, at a rate rapid enough to be attributed to a space-charge effect.

PERSPECTIVE ON THE FUTURE OF CONDITIONING

Chemical conditioning is only one of the alternatives open to the power industry in dealing with difficult fly ash. The technology of conditioning has been tried far more often on an experimental basis than it has been adopted for full-time continuous use. When permanent installations for conditioning have been made, usually they have been intended to upgrade the performance of an existing precipitator, not to ensure adequate performance by a new collector.

A major factor leading to this policy has been the unpopularity of conditioning with sulfur trioxide, mainly as a result of operating difficulties with injection systems for this agent and hazards associated with the compound. Another factor has been the frequent lack of success with ammonia and other flue-gas additives as alternative agents and the absence of an understanding of the mechanisms by which they operate.

Some of the needs that must be met if conditioning is to have a continuing or expanding application are: (1) development of new agents that are attractive on the basis of cost and convenience and (2) research on the mechanisms of action by various agents that will give a more dependable basis for predicting performance. Coupled with these developments, there must be a careful diagnosis of the specific problems in individual power plants that cause inefficient precipitation and a recognition of the fact that conditioning cannot rectify all difficulties. In other words, chemical companies, equipment vendors, and utility companies must all accept their responsibilities if, in the long range, conditioning is to serve the immediate interests of all concerned.

One final point to be made is that research must be done to ensure that conditioning has a favorable overall impact on the environment. Regulatory agencies will not, in the long run, tolerate conditioning as a means of reducing fly-ash emissions if it leads to other undesirable effects that can be avoided through the adoption of some other corrective measure.

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RAPPING REENTRAINMENT STUDIES

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ABSTRACT

The results of experimental investigations of rapping reentrainment in four electrostatic precipitators connected to coal-fired boilers and in a large pilot electrostatic precipitator are reported. The objective of the studies was quantification of rapping reentrainment in terms of the percentage and particle size distribution of the reentrained particulate.

Impactors and mass trains were used to measure separately rap and nonrap emissions. An extraction real time optical particle sizing system was also used to determine the percentage of rapping emissions in five particle size bands. This system also supplied data on the temporal variations of the emissions due to rapping. During the pilot-scale test and during one of the field tests, the rapping emissions were observed to depend on the time interval between raps. Rapping emissions decreased from 53% to 18% of total emissions for the pilot test as the time interval between raps was increased. The percentage contribution of rapping reentrainment to total emissions for all tests ranged from a high of 53% to a low of 6.5%. Data on the particle-size distribution of the rapping puffs indicated that the mass median diameters of the particles were on the order of 10-20 μm. The large particles emitted during the rapping puffs appeared to be agglomerates. Rapping emissions were observed during the pilot-scale test to have a significantly higher concentration in the lower half of the precipitator. One of the major causes for this appeared to be hopper "boil-up". It was photographically observed that most of the particulate dropped into the hoppers; then a portion rebounded and slowly escaped over the hopper baffles to exit the precipitator.

RAPPING REENTRAINMENT STUDIES

INTRODUCTION

Rapping reentrainment accounts for significant losses from many electrostatic precipitators. The losses occur when the collected dust is removed from precipitator collection plates by rapping or shaking. The dust is lost by direct reentrainment in the gas stream and by dispersion when the dust falls into the collection hoppers.

The rapping losses are in addition to losses from uncollected material and to losses from other sources of reentrainment. Other sources of reentrainment can include the direct scouring action of the gas on the collected dust, scouring during sparking, and sweepage of dust directly from the hoppers, caused by poor gas-flow conditions or by air inleakage to the hoppers.[1]

This paper discusses the results of experimental investigations of rapping reentrainment of fly ash in four electrostatic precipitators connected to coal-fired boilers,[2,3] and the results of an experimental investigation of rapping reentrainment of fly ash in a large pilot precipitator at FluiDyne Engineering Corporation's Rosemount Laboratory.[4]

The main objective of the studies was quantification of rapping reentrainment in terms of the percentage of total emissions and in terms of the particle size distribution of the reentrained particulate with the intent of providing data for inclusion of losses due to rapping reentrainment in a systems model of electrostatic precipitation. In addition, during the studies using the pilot precipitator, the basic mechanics of removal of dry dust by rapping and the variations in the removal mechanisms with changes in dust properties were investigated.

Background

The removal of particulate from a gas stream by a dry electrostatic precipitator involves four separate processes. The first is the precipitation of the particulate from the gas stream onto the collection plate. A Southern Research Institute report discusses the physical mechanisms involved and describes a computer model for the dust collection from the gas stream.[5] The second process is the removal of the particulate from the collection plate, and the third is the transfer of this particulate to a hopper. The fourth process is the removal of the collected particulate from the hopper to a disposal area.

In present-day precipitator applications, two approaches are prevalent with regard to the removal and transfer of the particulate from the collecting plates. One approach is to rap often and to provide maximum rapping acceleration to these plates during each rap in an attempt to minimize the thickness of the residual dust layer. The other approach is to vary the intensity and frequency of rapping in an attempt to minimize the quantity of material reentrained. A determination of the best rapping technique for a specific application depends on an understanding of the method by which dust is actually removed and transferred from the collection plates during a rapping sequence and of the effects of residual dust layers.

The mechanics of the dust removal process vary with the properties of the dust, precipitator operating conditions, and rapping parameters. Dust properties and precipitator operating conditions affect the adhesion and cohesion of the dust layer. The adhesion and cohesion of dust layers depend upon particle-to-particle forces. According to Tassicker,[6] the component forces are: London-van der Waals, triboelectric, capillary, surface dipole, and electric-field corona forces. These component forces are influenced by the following: particle diameter, porosity and compaction of the layer, complex dielectric constant, humidity in the gas, adsorbed surface dipolar molecules, work-function interfaces on the material, and the electric field and current density in the dust layer.

Theory of the Mechanics of Dust Removal. An elementary theory of dust removal which considers only the tensile strength (P) of the dust layer and the acceleration (a) normal to the plate has been developed by Tassicker.[7] The theory predicts that the dust layer is removed only when

$$a > \frac{P}{\delta \ell} = \frac{P}{M/A} \tag{1}$$

where δ is the bulk density, ℓ is the dust layer thickness, and M/A the mass per unit area. According to this equation, for a given dust thickness, the rapping intensity must be of sufficient magnitude to produce an acceleration greater than the ratio of the tensile strength of the ash layer and the mass per unit area. For a given acceleration, dust is removed only when

$$M/A > P/a; \tag{2}$$

that is, when the mass per unit area (dust surface density) is greater than the ratio of dust layer tensile strength to the normal plate acceleration. Since the mass per unit area depends on the dust layer thickness, which in turn is related to collection time between raps, the time interval between the raps is directly related to the efficiency of dust removal from the plates.

Experimental data obtained by Sproull[8] and by Penney and Klingler[9] show that the requirements for removal of a precipitated dust layer are in basic agreement with Tassicker's elementary theory for dust removal.

Emissions Due to Rapping. Emissions due to rapping and their dependence on rapping parameters have been reported previously by Sproull,[10] Plato,[11] Sanayev and Reshidov,[12] Schwartz and Lieberstein,[13] and Nichols, Spencer, and McCain.[2] Sproull found that reducing the intensities of the raps led to a reduction in rapping emissions. He also observed a vertical stratification of the emissions during rapping, with higher concentrations in the lower portion of the precipitator. The work by the others showed improvements in performance of full-scale precipitators when the time intervals between raps were increased.

Although these previous studies have added to our understanding of rapping reentrainment, they have not provided quantitative data on the amounts of emissions due to rapping reentrainment.

EXPERIMENTAL PROCEDURES

A complete characterization of rapping reentrainment requires the measurement of a large variety of variables. A block diagram of an experimental layout for our pilot rapping reentrainment study is shown in Figure 1. In addition to the data that is obtained with this arrangement, a complete characterization utilizes the precipitator design data.

The field experiments included a similar set of measurements to those made during the pilot studies. However, sampling viewing ports for photographing rapping emissions and for determining the vertical stratification of the rapping emissions were not available in the full scale units nor were load cells for measuring the quantity of fly ash collected on the collection plates. Hence these measurements were not included in the field tests.

The adhesion and cohesion properties of the collected dust layers were also not determined during each test since suitable field test equipment was not available. However, some information about the tensile strength of the collected ash layers was obtained during the pilot studies by using the elementary theory of adhesion given in the introduction, the load cell data, and plate acceleration data.

Quantification of rapping reentrainment in terms of the percentage of total emissions and in terms of the particle size distribution of the reentrained particulate required determining the amount of emissions with and without rapping reentrainment. To define the mechanisms by which the rapping losses occurred required time-resolved data on the particulate concentrations and size distributions across typical portions of the precipitator exit plane.

During the investigations the above requirements were met by using real-time systems and integrating systems. A real-time system based on an optical single-particle counter was developed which included a five-channel analog ratemeter, which provided parallel monitoring of the instantaneous concentrations of particles in five pre-selected size intervals ranging from 0.6 to 20 μm.

To conform to instrumental limitations on the total concentration of aerosol particles in the sample gas stream arriving at the sensor, the aerosol sample from the flue was diluted before measurement. Because of the very steep gradient in the size distribution, on a number basis, at the exit of a precipitator on a power boiler the diluter was made as a size-selective device which, under ideal conditions, dilutes small particles in the sample gas stream by fairly large factors while passing a relatively confined and undiluted stream of the lower concentrations of large particles directly to the particle sensor. Figure 2 illustrates the operating system for the particle dilution train. Because of the potentially large probe losses that can result from impaction and settling in the probe, the system was intended to be used only from below a duct with a minimum number of bends (one) between the sampling point and the particle sensing volume. The probe was oriented in the vertical direction so that settling was in the direction of the probe axis and consequently did not result in a loss of particles.

For those circumstances in which it was not possible to sample from below the duct, a second sample extraction system was constructed. These samples were removed at high flow rates, 0.002-0.02 m^3/sec (5-40 cfm) through a large bore probe (4 cm diameter) and conveyed to a suitable location beside or on the top of the duct, at which point a secondary sample was extracted into the diluter and counter as illustrated in Figure 3. This sampling method provided information on relative concentrations of particles of various sizes during and between puffs, but did not provide quantitative concentration data because of the uncertainties in the probe losses and in the degree to which the secondary sample represents the average concentration in the high flow rate probe.

Integrated measurements of rapping emissions were made with separate sets of inertial impactors and mass trains during the pilot studies to independently quantify rapping reentrainment losses and non-rapping reentrainment losses. One set of impactors and mass trains measured losses between raps and the other set measured rapping losses. The dust feed was turned off during rapping. During two of the field tests, losses were measured with separate sets of impactors and mass trains during alternating periods with rapping and without rapping. The weighted time average of the measurements was compared with data obtained during normal operation to determine if losses were affected by use of the non-rapping intervals. During the other two field tests, measurements were made during non-rapping periods and compared with measurements during either normal operation or operation with a modified rapping frequency.

RESULTS OF EXPERIMENTAL STUDIES

In the tests on the five electrostatic precipitators, the following data were obtained: the percent of total emissions due to rapping reentrainment, the percent of emissions for a given particle size due to rapping, the particle size distribution of the particulate emitted due to rapping, and the effects of rapping reentrainment on fractional collection efficiencies. Selected examples of the test results are contained in this paper.

The percentage contribution of rapping reentrainment to total emissions at each of the test locations and for various test conditions are tabulated in Table 1. The percentage contributions ranged from a high of 53% to a low of 6.5%.

The percentages of emissions for a specific particle size that are due to rapping are given in Figure 4. This data indicates that rapping reentrainment accounts for only 2-5% of the emissions of 1 μm diameter particles, and that the percentage of emissions due to rapping increases significantly with increasing particle size, ranging from 24% to 48% for 4 μm diameter particles and from 50% to 88% for 12 μm diameter particles.

Averaged particle size distributions for rapping puffs are plotted in Figures 5 and 6. The particle size distribution data shows that the rapping puffs consisted of relatively large particles when compared with non-rapping emissions.

A selected example of the effect of rapping reentrainment on precipitator fractional efficiency is shown in Figure 7. Fractional efficiencies for alternating periods with the rappers off and with the rappers on are given along with fractional efficiencies determined during normal rapper operation. These data show that collection efficiency is relatively independent of rapper operation for particles with diameters from 0.1 μm to 2 μm with the possible exception in the region of 0.3 to 8 μm in which there is some ambiguity in the data. Above 2 μm, rapping significantly reduced collection efficiency.

Discussion of Experimental Results

The percentage contribution of rapping reentrainment to total emissions varied significantly for changes in test conditions at the same location and for the different precipitators. The lowest rapping emissions were apparently obtained for a precipitator with high plate accelerations, i.e., accelerations greater than 150 G's (1.5×10^5 cm/sec^2) and with a rap interval of 30 minutes. There are several possibilities for the apparent low rapping emissions at this location that are of interest and which merit further investigation. One is the use of large and relatively deep hoppers (depth 6 meters). It is suspected that these aided in reducing hopper "boil-up", which is discussed in the section on reentrainment mechanisms. Another possibility is the effect of having a relatively fine inlet dust. There also is the possibility that other sources of reentrainment may have contributed to non-rap emissions, resulting in the apparently low rapping emissions.

The highest percentage rapping emissions were obtained during the pilot studies with a rapping interval of 12 minutes and with normal plate accelerations on the order of 11 G's (1×10^{11} cm/sec^2). The rapping emissions during the pilot study (test location number 2) decreased with increasing time between raps. The effect on overall efficiency is shown in Figure 8. The percentage of the collected dust removed from the precipitator plates also increased with increased time between raps, as shown in Figure 9.

The collecting of a larger mass per unit area (dust surface density) produces the two effects mentioned above. According to the elementary theory of dust removal, the product of the normal plate acceleration and dust surface density has to be greater than the tensile strength of the layer, typically 0.5×10^3 to 3×10^3 dynes/cm^2, if the dust layer is to be removed. This theory indicates that dust removal efficiency should increase with increased time between raps, as illustrated by the data shown in Figure 9. Laboratory measurements by Sproull[8] produced similar data. Figure 9 also illustrates the build-up of a residual dust layer that was not removed with normal plate accelerations on the order of 11 G's. There are several possible causes for the development of the residual layer. For one, the dust layer directly in contact with the collection plates has a much higher tensile strength than the remainder of the layer. Estimates for removal of the layer called for accelerations greater than 10^3 G's (9.8×10^5 cm/sec^2).[7] Consolidation of the dust that remains on the plate after a rap also aids in producing residual layers. The vibrations during a rap can have the effect of compacting the dust layer if it is not removed making it more difficult to remove. A third possible cause of the residual layer is the removal of patches of dust only from selected locations on the collection plates where the removal criteria are met. Dust can be removed from one location during one rap and from another location on the next rap due to changes in distribution of the dust surface density. This results in a nonuniform dust layer and the presence of a residual layer. This is often the result of nonuniform plate accelerations. At one location where plate accelerations are on the order of only 4 to 5 G's ($3.4\text{-}4.9 \times 10^3$ cm/sec^2) residual dust layers as thick as 2 cm were observed in the vicinity of plate baffles where the plate accelerations are dampened. Between the baffles, the residual layers were only 1 to 2 mm thick.

The decrease in rapping emissions observed during some of the tests with an increase in the time between raps or correspondingly an increase in dust surface density appears to be the result of a shift in the size distribution of the reentrained particulate to larger particle sizes. Thicker dust layers appear to produce larger particles. In this case the large particles that are emitted are not the primary particles that entered the precipitator but agglomerates consisting of smaller particles. Inspection of impactor substrates after collecting dust at the outlet of test locations number 2 and 3 indicated that the majority of the large particles in the rapping puffs were agglomerates. The shift to larger particles in the rapping puffs observed during the pilot studies as the time between raps was increased is illustrated by the shift of the cumulative percent particle size distribution of the rapping puffs shown in Figure 5.

This effect was previously observed by Plato,[11] who found that with a dust layer surface density greater than 1 kg/m^2 dust was removed in cakes and for values less than this it formed clouds. Producing agglomerates instead of individual particles when the plates are rapped reduces the quantity of material that escapes from the precipitator due to rapping. This is because the large agglomerates are recollected faster by gravitational settling and by electrostatic precipitation than discrete single particles or smaller agglomerates.

The above provides an explanation for the results that are given in Figures 8 and 10. Sanayev and Reshidov have also used the above explanation for the decrease in the percentage of dust collected on the plates that is emitted due to rapping with an increase in dust surface density. A comparison of results obtained during the pilot studies[4] and by Sanayev and Reshidov[12] is shown in Figure 11.

Both sets of data indicate that for the conditions used during the tests, minimum reentrainment loss occurred with a dust surface density on the order of 1 kg/m^2 or larger. The Sanayev and Reshidov data also show the expected dependence on gas velocity. The results given in Figure 11 do not include the effects of collecting thicker layers on overall performance.

The 50 to 88% emissions due to rapping reentrainment observed for large particles (>10 μm) which are shown in Figure 4 can be explained either by a continuing transfer of collected large particles from the inlet to the outlet by rapping or by the above mentioned agglomeration of the collected particles. Both effects probably occur, with agglomeration being the dominant effect, as indicated by the agglomerates observed on the upper stage of inertial impactors.

During collection by an electrostatic precipitator, the mass median diameter of the fly ash from a coal-fired boiler typically decreases from the inlet of the precipitator to the outlet. This is illustrated by a shift in the particle size distributions of the inlet and outlet dust concentrations. In

Figure 6, the dm/dlogD particle size distributions are plotted for the inlet dust at test location number 4 and for the outlet dust with the rappers off and with the rappers on. The expected shift in particle size which can be theoretically predicted for the inlet and outlet dust concentrations with the rappers off is clearly evident. When the plate rappers are on, the mass median diameter of the particle size distribution shifts back towards larger size particles as shown by the dm/dlogD plot in Figure 6 with the rappers on and by the effective average dm/dlogD for the rapping puffs. The large particle size of the particulate in the rapping puffs is also illustrated by the data given in Figure 5. The rapping emission percentages in Figure 4 reflect the shift in the outlet particle size distribution produced by plate rapping. At locations such as at test location number 3 where the precipitation is very large and preceded by a mechanical collector almost all of the large particles ($\gtrsim$10 µm) are collected before reaching the outlet. Thus the percentage of emission due to rapping can approach 100% for the large particles.

Rapping reentrainment has the effect of transferring particles from particle size intervals consisting of small single particles to particle size intervals consisting of larger particles.

At test location number 4, data were obtained at normal operating current densities (see Table 1) and at one-half the normal operating current density. This resulted not only in a decrease in efficiency between raps but also, as shown by the data given in Figure 4, in an increase in the percentage contribution of rapping reentrainment for large particles. This implies that a change in particulate properties such as an increase in resistivity that reduces operating current density can result not only in a decrease in efficiency between raps but also an increase in the percentage emissions due to rapping reentrainment. However, since the forces holding the dust layer together theoretically increase with resistivity, the effects may not be the same as those obtained during the test at location number 4, where dust resistivity was held constant.

Reentrainment Mechanisms

The amount of rapping reentrainment that occurs depends on the processes by which the dust is reentrained into the gas stream. Sproull[10] has suggested that the rapping processes occur in a sequence in which the collected dust is removed by what is termed a slip-stick process. The concept is that the dust layer builds up to a sufficient thickness so that the acceleration of the plate causes a shearing force between the dust layer and the plate. The dust then slides along the surface of the plate until it is recollected on the plate. The process is repeated with each successive rap until the dust finally falls into the hopper.

Another concept of rapping is that the collected dust is dislodged in large agglomerates or clumps and falls directly into the hopper without being recollected.

Neither concept adequately explains the rapping process as observed in the FluiDyne pilot precipitator or in the Southern Research Institute small scale precipitator. Motion pictures (32 frames/sec) of the removal of a dust layer (2-3 mm thick) by rapping in the SRI pilot unit show the dust layer fracturing along lines of discontinuity in the dust surface. The resulting fractured sheet of dust starts to fall as separate sheets which break up as they encounter other falling sheets and patches of unremoved dust. The dust appears to fall without being recollected and to become turbulently mixed as it falls. The motion pictures show the majority of the dust dropping into the hoppers from which a portion boils up and becomes reentrained into the gas stream. Motion pictures taken in the large pilot precipitator at the FluiDyne Engineering Rosemount Laboratory showed similar behavior.

Measurements of the vertical distribution of the rapping loss at the exit of the FluiDyne unit indicated that 82% of the rapping emission occurred in the lower 32% of the precipitator. This effect, evidently due to hopper "boil-up" combined with gravitational settling of the reentrained particulate produced a significant vertical stratification of the reentrained emissions from the large pilot precipitator as shown in Figure 12.

Rapping puffs observed in the lower portion of the precipitator occurred in two bursts for both upstream and downstream raps as shown in Figure 13. The first burst lasted 2-4 seconds. This burst was interpreted as being the result of particulate reentrained directly in the gas stream and being carried out of the precipitator at the velocity of the gas through the unit. The longer lasting second burst, which for the larger particles was a series of puffs, can be interpreted as resulting from hopper "boil-up". These data indicate that hopper "boil-up" contributes significantly to rapping reentrainment emissions.

EFFECT OF RAPPING REENTRAINMENT ON COLLECTION EFFICIENCY, PENETRATION, AND SIZING OF ELECTROSTATIC PRECIPITATORS

Rapping reentrainment can have a considerable effect on collection efficiency, penetration, and sizing of electrostatic precipitators. In this section two approaches to estimating the effects of rapping reentrainment are discussed. The examples are based on the results of the FluiDyne pilot tests and computer calculations by Gooch, et al.[5] The results of the example calculations are tabulated in Tables 2 and 3.

These examples show that the percentage increase in collection efficiency obtained with no rapping reentrainment decreases with increasing efficiency of the precipitator. However, due to the exponential dependence of precipitator efficiency on particle migration velocity, plate collection area, and gas volume flow rate, the effect of rapping reentrainment on the size of a precipitator required for a given collection efficiency is substantial. Estimates of the effect of rapping reentrainment on precipitator size are tabulated in Tables 2 and 3 for various conditions. Two approaches were used to generate the tabulated values. One approach was to assume that the percentage penetration due to rapping was independent of the size of the unit. This assumption, although not strictly true, may fairly represent real situations. According to the data obtained during the pilot experiments, the rapping puff concentration has a significant vertical gradient and the rapping puff consists of large particles. Thus, in a multi-section precipitator the rapping puffs from upstream sections may be recollected in the lower portion of a downstream section and later transferred from this lower position into the hoppers without making any significant contribution to outlet emissions. In this case, outlet fields would contribute the largest portion of the rapping emission, and the percentage emission due to rapping should be relatively independent of the size of the unit. Estimates based on this approach are tabulated in Table 2.

The second procedure was developed by Francis.[14] This approach assumes that the fraction of collected dust per section reentrained and emitted is the same for every section, that the between-rap sectional collection efficiency is constant from one section to another, and that the recollection efficiency for the rapping puffs is the same as the between rap sectional collection efficiency. According to Francis' procedure, the penetration of any downstream section is equal to the penetration of the first section (P_1) raised to the power of the number of preceding sections plus one. Thus, the penetration of the N section (P_N) is given by:

$$P_N = (P_1)^N \tag{3}$$

Using this equation, the collection efficiencies of a precipitator having 1, 2, 3, or 4 sections with a between-rap sectional collection efficiency of 94.8% were calculated for percent penetrations of the first section due to rapping reentrainments of 52%, 30%, and 15%. These percentages correspond to 5.9%, 2.3%, and 0.97% of the collected material per section being reentrained and emitted from a section. These values were chosen to correspond approximately with those obtained during the FluiDyne experiments. The results of the calculations based on this procedure are tabulated in Table 3. According to this procedure, for the conditions corresponding to the worst rapping reentrainment case at FluiDyne, the contribution of rapping reentrainment to total emissions would be greater than 90% for a three or four section precipitator with 94.8% sectional collection efficiency. For a case in which reentrainment accounts for only 15% of the emissions for the first section, a situation slightly better than the best obtained during the FluiDyne experiments, rapping reentrainment, according to Francis' procedure, would account for 39% of emissions for a 3-section precipitator and 50% of emissions for a 4-section precipitator, if the precipitator had a sectional efficiency of 94.8%. These calculations indicate that, as the size of the precipitator increases, the percentage contribution of rapping reentrainment to total emissions increases. As previously discussed, this may not happen. Investigations of rapping reentrainment in multi-section precipitators do not appear to show a general increase in the percentage contribution of rapping reentrainment to total emission with precipitator size except for large particles. To obtain an indication of the effect of rapping reentrainment on the sizing of a precipitator, estimates of the theoretical specific collection area needed to obtain a given collection efficiency for a particular set of conditions were obtained from the model developed by Gooch, et al.[5] For the example chosen, a typical particle size distribution for an ash from a coal-fired boiler plant was selected and a precipitator operating current density of 20 nA/cm^2 was assumed (see the report of Gooch, et al,[5] for further details).

The theoretical specific collection area (SCA) required to obtain a desired collection efficiency with rapping reentrainment was compared with a theoretical SCA that would give the same efficiency if there were no rapping emissions. This gave an estimate of the increase in SCA needed to counterbalance the rapping reentrainment emissions. The theoretical SCA needed to obtain the desired collection efficiency with rapping reentrainment was determined by first computing the collection efficiency that would be obtained if there were no rapping reentrainment, by assuming that either the percentage of emissions due to rapping reentrainment or the fraction of collected dust reentrained and emitted from a section due to rapping were known. Once this collection efficiency was calculated and assuming that there were no other nonidealities beside rapping reentrainment, the required theoretical SCA was obtained from data given by Gooch, et al. Then, again using the data given by Gooch, et al, the SCA was determined that would give the desired efficiency if there were no nonidealities such as rapping reentrainment. For the particular cases considered, the increase in required SCA due to the rapping reentrainment ranged from 6% to 84% (see Tables 2 and 3). Thus, although the percentage decrease in collection efficiency due to rapping reentrainment is relatively small, the effect on the size required for a given collection efficiency can be significant. The data obtained from the FluiDyne pilot experiments indicates that even moderate changes in rapping parameters can significantly affect the size of a precipitator for a given collection efficiency.

SUMMARY

Experimental investigations of rapping reentrainment were conducted utilizing a large pilot electrostatic precipitator and several electrostatic precipitators installed at coal-fired power plants. The objectives of the experimental study included a fundamental study of the mechanics of removal of dry fly ash by rapping and the quantification of rapping reentrainment in terms of the percentage and particle size distribution of the reentrained fly ash.

During the experiments the percentage of fly ash removed from the plates was observed to depend on the mass per unit area of ash collected on the plates, as predicted by theory. The build-up of residual dust layers was also observed.

During the tests the contribution of rapping reentrainment to the total emissions ranged from 6.5 to 53%, depending on the rapping conditions and the precipitator design. A significant decrease in rapping emissions was obtained by increasing the time interval between raps. This decrease was due to the resulting larger mass per unit area collected on the plates when they were rapped. Particle size distribution measurements showed that the mass median diameter of the particles emitted during the raps increased with increased time between raps. As would be expected, this produced lower overall emissions. The increase in the size of the particles emitted during rapping was ascribed to an increase in the extent of agglomeration of the particulate removed from the precipitator plates with the corresponding increase in the mass per unit area collected on the plates (thicker dust layers) before the plates were rapped.

A major portion of the reentrained material was observed to result from hopper "boil-up." A small portion of the dust was observed to pass out a pilot precipitator in a short burst at the velocity of the gas passing through the unit, while the remaining portion of the material was observed to fall into the hoppers, then to rebound, and finally to escape slowly over the baffles and out of the precipitator. This produced a significant vertical gradient in the concentration of the dust emitted from the precipitator due to rapping reentrainment.

The fractional collection efficiencies of the precipitators both excluding and including rapping reentrainment emissions were determined. Rapping reentrainment decreased the fractional collection efficiencies for particles larger than 2 μm. Below 2 μm, rapping reentrainment did not appear to contribute significantly to overall emissions.

Two approaches using simplified assumptions were used to estimate the effect of rapping reentrainment on the size of a precipitator. The first assumed that a fixed percentage of emissions was emitted from a precipitator independent of the size of the unit, and the second assumed that the same percentage of collected material was reentrained and emitted from each section and that the recollection efficiency for the reentrained material for all sections was the same as the efficiency between raps, with all sections having the same efficiency. The estimates based on the above approaches showed that the increase in precipitator size needed to recover the rapping reentrainment emissions can account for 6% to greater than 80% of the size of the unit for the cases considered.

The parameters studied during the experiments were somewhat limited. A data bank for detailed modelling of rapping should include additional data on rapping emissions as a function of plate accelerations and gas velocity.

The pilot experiments also indicated a need for a study of the parameters affecting hopper "boil-up" and the development of methods for reducing it. Additional studies on the effects of residual dust layers for high resistivity dust and studies of dust cohesive properties are also needed to aid in determining optimum rapping conditions.

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Table 1. PERCENTAGE CONTRIBUTION OF RAPPING REENTRAINMENT TO TOTAL EMISSIONS

Test Location	Comments	Type Rapper	Plate Acceleration G's	Rap Intervals Min.	Gas Velocity m/sec	Avg. Plate Current Densities** mA/cm^2	Gas Temp °C	SCA*, $\frac{m^2}{m^3/sec}$	Total Penetration %	Penetration Due to Rapping Reentrainment, %
			x,y,x axis							
1		Vibrator	3.7, 1.6, 2.9	4, 6.5, 6.5		7.6, 10.4, 15.3, 13.9	137		32.1	46
		Vibrator	3.5, 1.8, 2.6	8, 13, 13		11.2, 11.2, 16.7, 12.5	137		25.8	33
	SO_3 injection	Vibrator		4, 6.5, 6.5		11.2, 11.2, 16.7, 12.5	132		6.2	48
		no rap							17.4	
	SO_3 injection	no rap							3.2	
2	pilot test	Drop hammer	11, 16, 15	12	0.87	23.3	122	33	11.4	53
	pilot test	Drop hammer		32					7.6	32
	pilot test	Drop hammer		52					6.1	18
	pilot test	Drop hammer		150					6.9	25
	pilot test	no rap							5.2	
3		Drop hammer		10, 10, 20, 20	1.52	8.4, 9.1, 13.2	154	110	0.85	30
		two plates		60, 60		15.1, 12.3				
4		Drop hammer		10, 20, 60	1.25	11.1, 17.6, 22.7	157	48	0.40	35
		two plates				4.6, 7.8, 11.3	157		0.95	38
5		Rotating Drop hammer		6, 6, 12, 12	1.83	14, 24.6, 33.0, 47.7	162	52	0.19	6.5

*SCA - specific collection area

**for each electrical field

Table 2. EFFECT OF RAPPING REENTRAINMENT ON EFFICIENCY FOR RAPPING LOSSES ACCOUNTING FOR 20%, 30%, 40% AND 50% OF EMISSIONS FOR VARIOUS OVERALL EFFICIENCIES
(first approach)

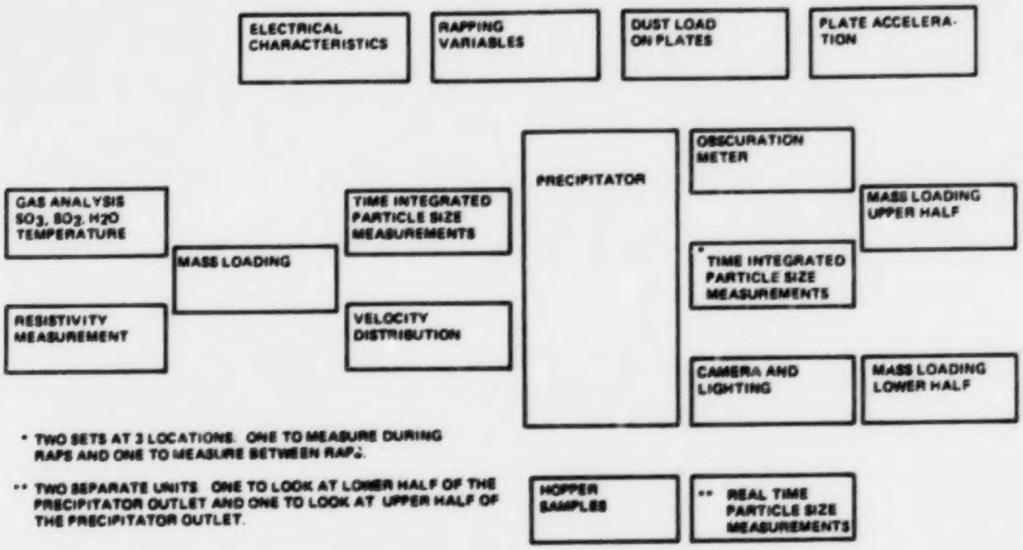
Assumed percent of total penetration due to rapping reentrainment	Penetration: Due to rapping reentrainment, %	Penetration: Without rapping reentrainment, %	Penetration: Total %	Assumed Overall Efficiency η %	Efficiency without rapping reentrainment, η_R, %	Increase in η without rapping reentrainment, %	SCA* required: For η without rapping reentrainment, $m^2/m^3/sec$	SCA* required: For η without rapping reentrainment, $(ft^2/1000\ ft^3/sec)$	SCA* required: For η_R or for η with rapping reentrainment, $m^2/m^3/sec$	SCA* required: For η_R or for η with rapping reentrainment, $(ft^2/1000\ ft^3/sec)$	Additional SCA required because of reentrainment $m^2/m^3/sec$	Additional SCA required because of reentrainment $(ft^2/1000\ ft^3/sec)$	Additional SCA, %
50	3.75	3.75	7.50	92.50	96.25	3.90			16.2	(82)			
	2.25	2.25	4.50	95.50	97.75	2.31			21.3	(108)			
	0.75	0.75	1.50	98.50	99.25	0.76	26.2	(133)	35.5	(180)	9.3	(47)	35.3
	0.25	0.25	0.50	99.50	99.75	0.25	41.8	(212)	53.2	(270)	11.4	(58)	27.4
	0.05	0.05	0.10	99.90	99.95	0.05	68.9	(350)	81.7	(415)	12.8	(65)	18.6
40	3.00	4.50	7.50	92.50	95.50	3.24							
	1.80	2.70	4.50	95.50	97.30	1.88			18.7	(95)			
	0.60	0.90	1.50	98.50	99.10	0.61	26.2	(133)	32.9	(167)	6.7	(34)	25.6
	0.20	0.60	0.50	99.50	99.70	0.20	41.8	(212)	50.2	(255)	8.5	(43)	20.3
	0.04	0.06	0.10	99.90	99.94	0.04	68.9	(350)	78.8	(400)	9.8	(50)	14.3
30	2.25	5.25	7.50	92.50	94.75	2.43							
	1.35	3.15	4.50	95.50	96.85	1.41							
	0.45	1.05	1.50	98.50	98.95	0.46	26.2	(133)	30.5	(155)	4.5	(23)	16.5
	0.15	0.35	0.50	99.50	99.65	0.15	41.8	(212)	47.3	(240)	5.5	(28)	13.2
	0.03	0.07	0.10	99.90	99.93	0.03	68.9	(350)	75.8	(385)	6.9	(35)	10.0
20	1.50	6.00	7.50	92.50	94.00	1.62							
	0.90	4.60	4.50	95.50	96.40	0.94							
	0.30	1.20	1.50	98.50	98.80	0.30	26.2	(133)	29.2	(148)	3.0	(15)	11.3
	0.40	0.10	0.50	99.50	99.90	0.10	41.8	(212)	68.9	(350)	3.5	(18)	8.5
	0.08	0.02	0.10	99.90	99.92	0.02	68.9	(350)	73.3	(372)	4.3	(22)	6.3

*Specific Collection Area (SCA) based on data given by Gooch, et al.[5] (Figure 26 computed performance curves at 20 nA/cm^2).

Table 3. EFFECT OF RAPPING REENTRAINMENT ON PERFORMANCE FOR VARIOUS REENTRAINMENT PERCENTAGES AS A FUNCTION OF NUMBER OF SECTIONS AND A CONSTANT COLLECTION EFFICIENCY PER SECTION.

		Penetration							SCA* Needed						
									Without rapping reentrainment to obtain η		With rapping reentrainment		Additional SCA		
Assumed percent of collected material per section reentrained	Number of Sections	Due to rapping reentrainment, %	Without rapping reentrainment, %	Total %	Percent of penetration due to rapping reentrainment, %	Efficiency η, %	Assumed efficiency η_R without rapping reentrainment, %	Increase in η with no rapping reentrainment, %	$m^2/m^3/sec$	$(ft^2/1000\ ft^3/sec)$	$m^2/m^3/sec$	$(ft^3/1000\ ft^3/sec)$	$m^2/m^3/sec$	$(ft^2/1000\ ft^3/sec)$	Additional SCA, %
5.9	1	6.2	5.2	11.4	52	88.6	94.8	6.54							
	2	1.02	0.27	1.29	79	98.7	99.73	1.03	28.0	(142)	51.6	(262)	23.6	(120)	84
	3	0.136	0.014	0.150	91	99.85	99.986	0.14	62.1	(315)	105.4	(535)	43.3	(220)	69
	4	0.0193	0.0007	0.020	97	99.98	99.9993	0.02	98.5	(500)	108.3	(550)			
2.3	1	2.23	5.20	7.43	30	92.57	94.80	2.35							
	2	0.28	0.27	0.55	51	99.45	99.73	0.28	40.4	(205)	51.6	(262)	11.2	(57)	28
	3	0.027	0.014	0.041	66	99.959	99.986	0.03	83.7	(425)	105.4	(535)	21.7	(110)	25
	4	0.0023	0.0007	0.0030	77	99.997	99.9993	0.002	108.5	(550)	108.3	(550)			
0.97	1	0.90	5.20	6.1	15	93.9	94.8	0.95							
	2	0.10	0.27	0.37	27	99.63	99.73	0.10	46.3	(235)	51.6	(262)	5.3	(27)	11
	3	0.009	0.014	0.023	39	99.97	99.986	0.016	91.0	(462)	105.4	(535)	14.4	(73)	16
	4	0.0007	0.0007	0.0014	5C	99.9986	99.9993	0.001	108.3	(550)	108.3	(550)			

*Specific Collection Area (SCA) based on data given by Gooch, et al.[5]
(Figure 26 computed performance curves at 20 nA/cm^2).



* TWO SETS AT 3 LOCATIONS. ONE TO MEASURE DURING RAPS AND ONE TO MEASURE BETWEEN RAPS.

** TWO SEPARATE UNITS. ONE TO LOOK AT LOWER HALF OF THE PRECIPITATOR OUTLET AND ONE TO LOOK AT UPPER HALF OF THE PRECIPITATOR OUTLET.

Figure 1. Block diagram of experimental layout for a rapping reentrainment study.

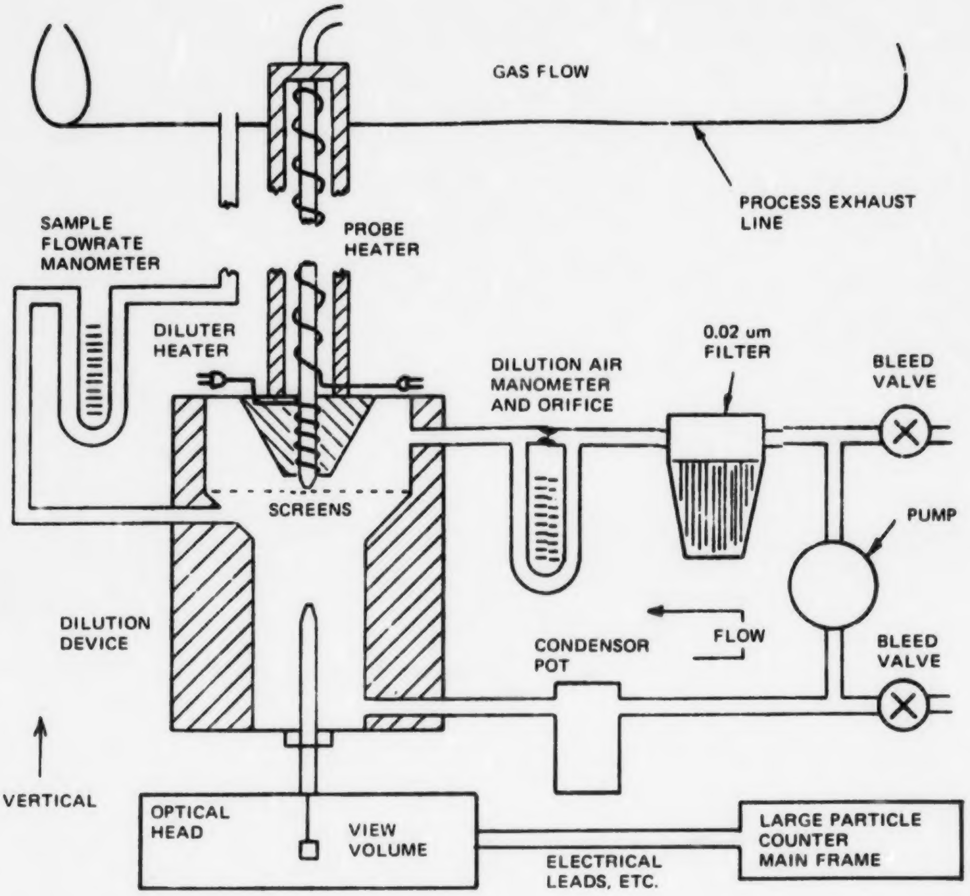


Figure 2. Schematic of diluter for real time particle sizing system.

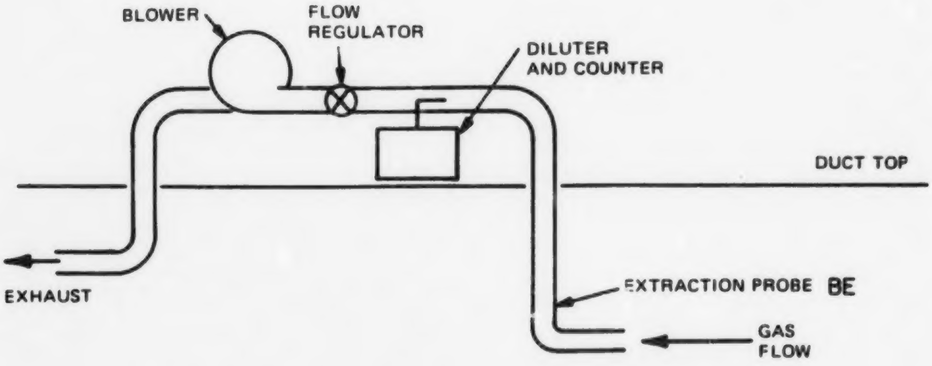


Figure 3. Extractive sampling system for real time system

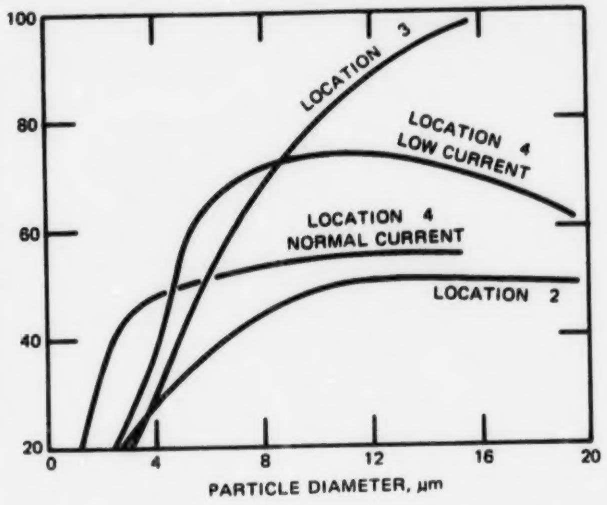


Figure 4. Percent emissions due to rapping for particles with diameters at l to 20 µm.

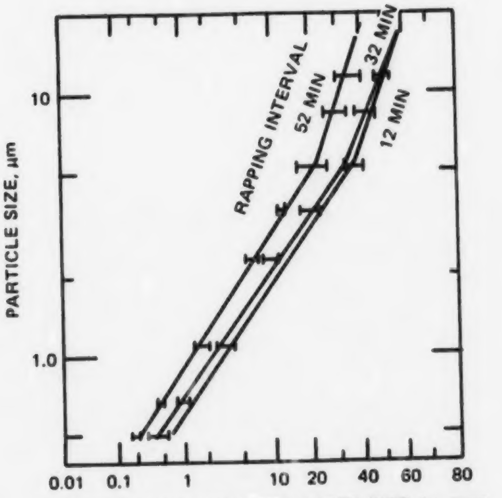


Figure 5. Cumulative percent distribution for rapping puffs, rapping intervals of 12, 32, and 52 minutes, pilot test.

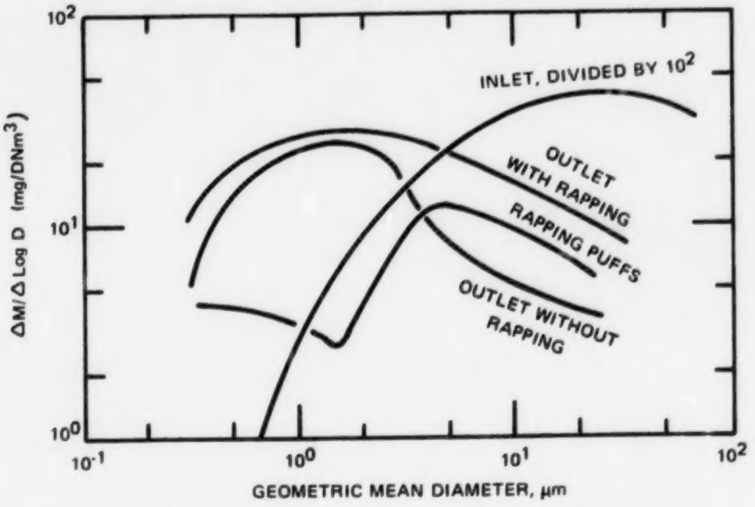


Figure 6. Particle size distributions for inlet particulate, outlet particulate with rapping, outlet particulate without rapping, and average effective distribution for rapping puffs.

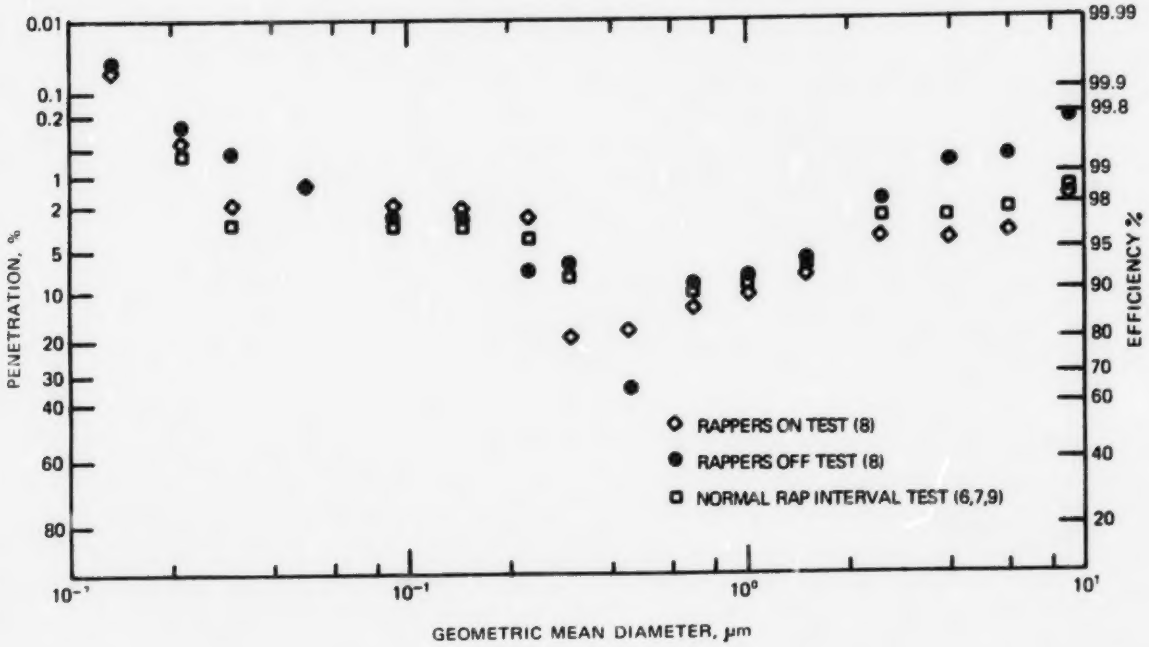


Figure 7. Fractional efficiencies as function of particle diameter for half normal current density test. Test Location Number 3.

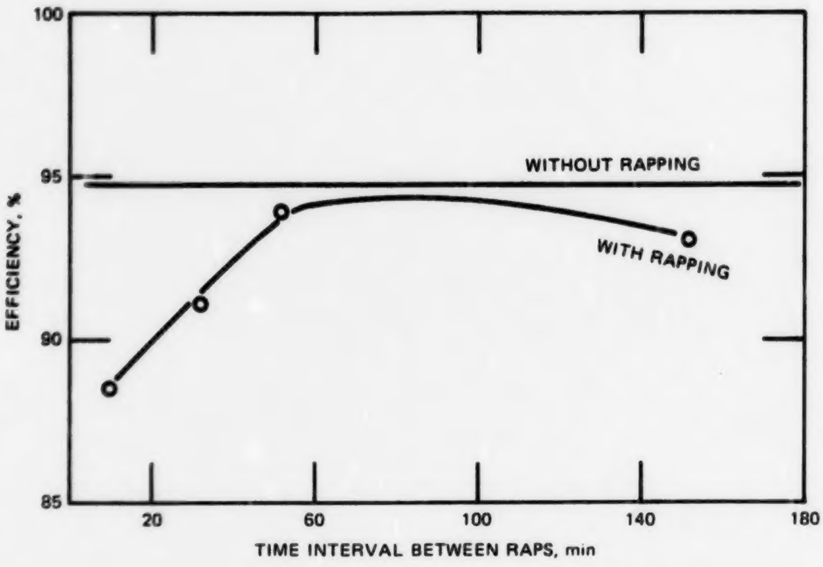


Figure 8. Average efficiencies for FluiDyne pilot precipitator for various rapping intervals.

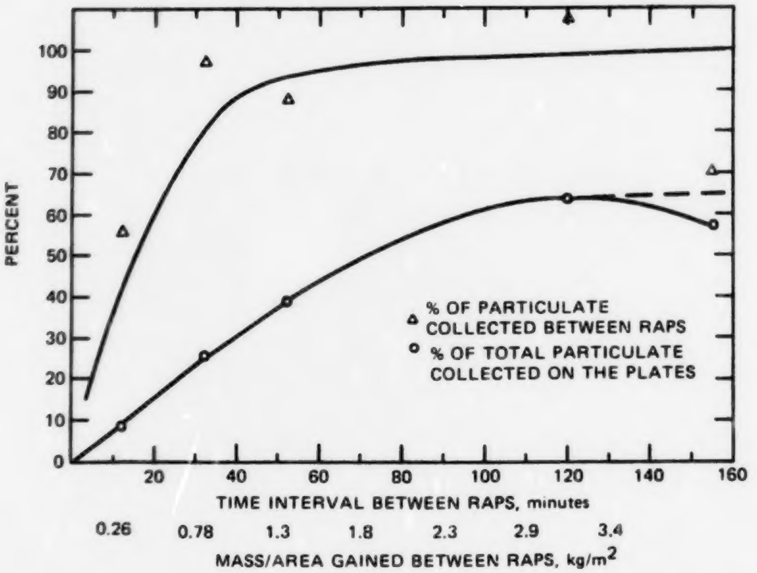


Figure 9. Dust removal efficiency versus the time interval between raps.

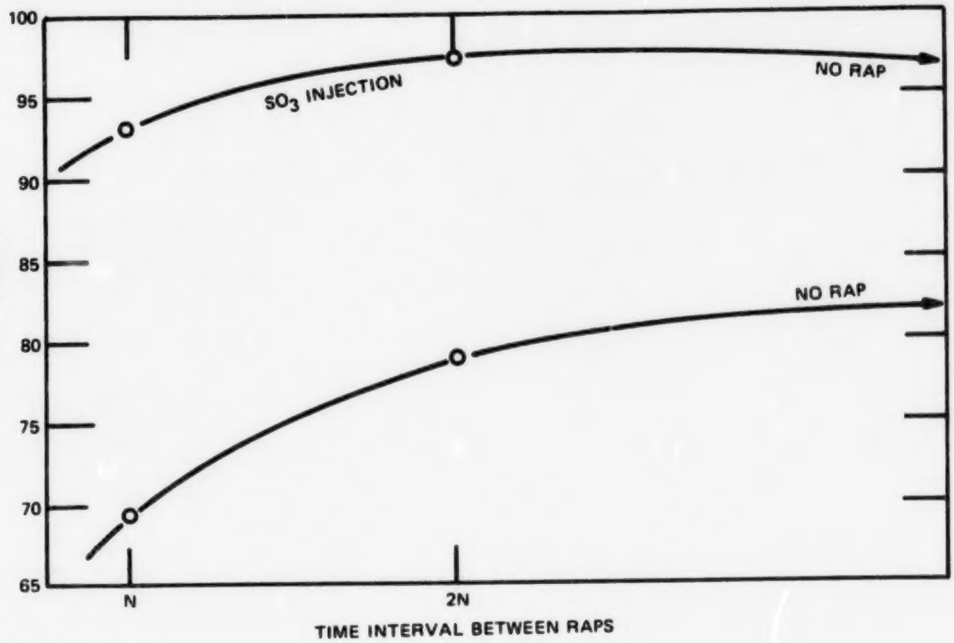


Figure 10. Precipitator efficiency at Test Location Number 1 as a function of the rapping interval.

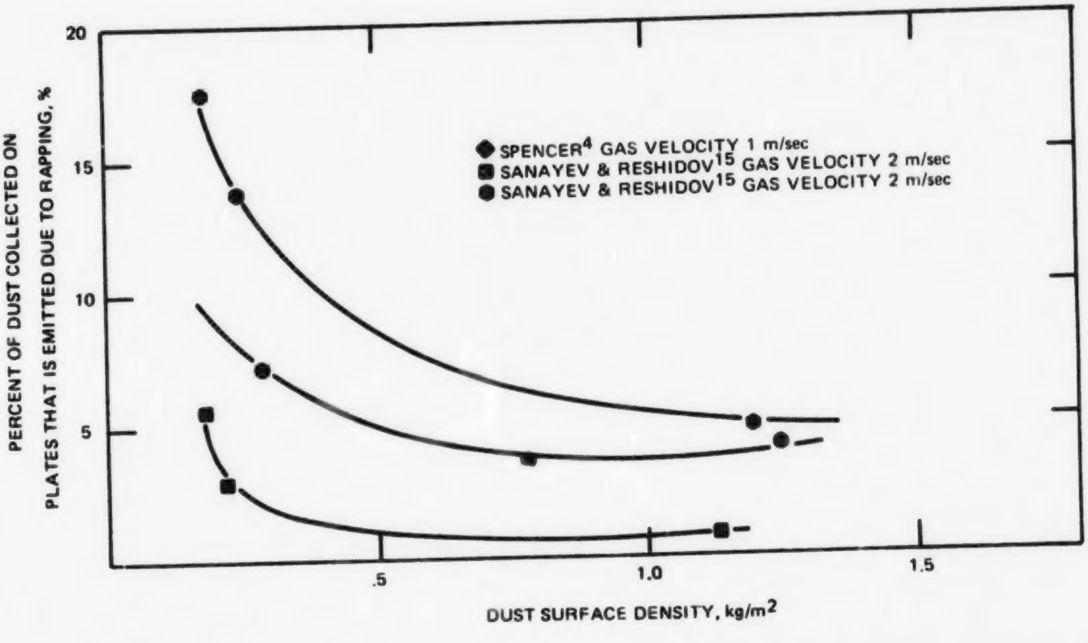


Figure 11. Percent of dust collected on precipitator plates emitted from pilot precipitators due to rapping.

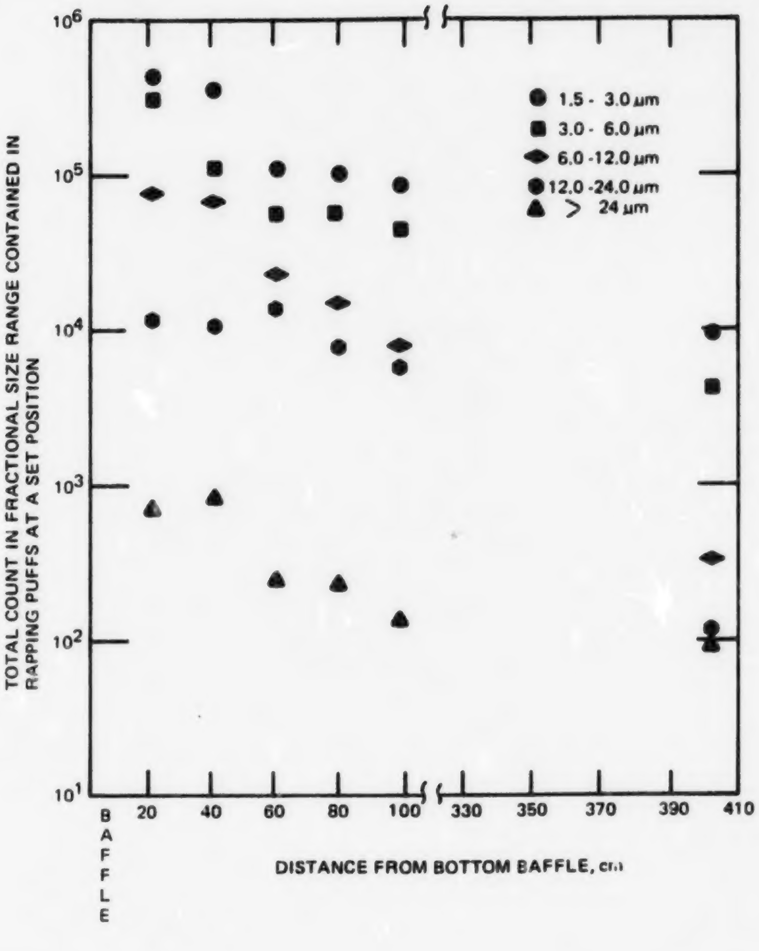


Figure 12. Spatial distribution of particles in rapping puff.

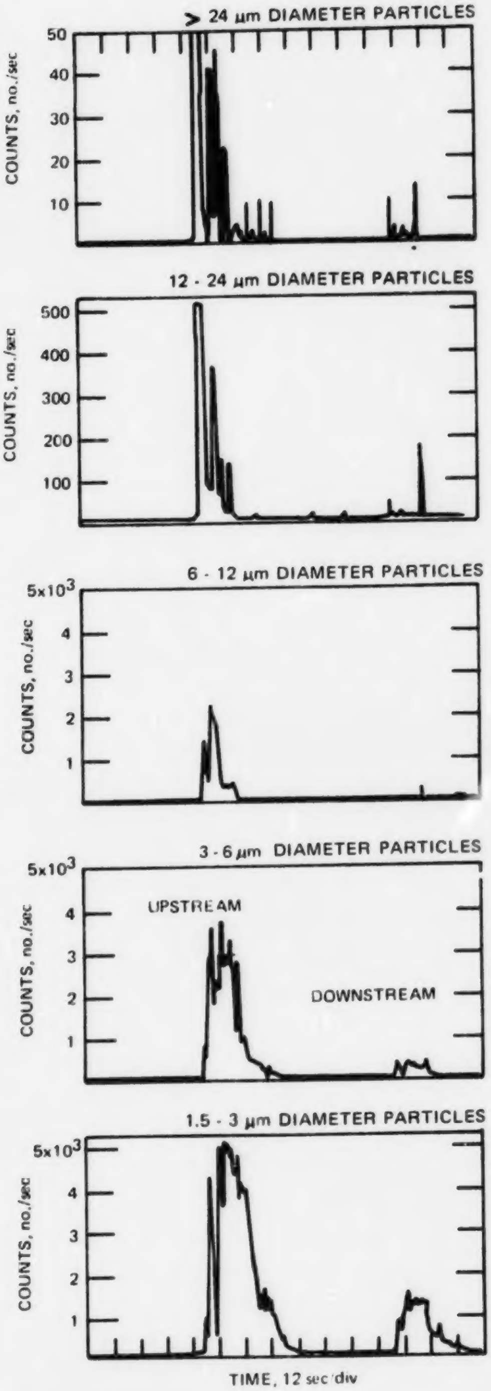


Figure 13. Rapping puffs at the exit plane of the pilot precipitator, upstream and downstream raps.

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FABRIC FILTRATION PERFORMANCE MODEL

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ABSTRACT

The goals of this program were to formulate and develop mathematical methods for predicting and analyzing fabric filter drag (pressure drop per unit face velocity) and fabric filter collection efficiency as a function of particle size. The literature with respect to filtration is vast, but the efforts to model *fabric* filtration have been few. Most attempts have used a linear approximation to the increase of fabric drag with added weight per area, and those who attempted to model collection efficiency did so with models based upon arrays of cylinders. The woven fabric structure is much more like an array of pores (before the cake is formed) and this representation has given rise to a new, nonlinear model of fabric drag and a new model of fabric collection efficiency, both presented here. The nonlinear drag model fits the experimental data much better than does the linear. The efficiency model appears to give results which are both reasonable and informative. These models have been combined to form a preliminary computer program for an entire multi-bag filtration installation, and results of such modeling are presented for both flow and efficiency variables. Also presented are design equations which are readily adaptable for a programmable calculator. Finally, we indicate several areas in which further investigation seems warranted in analyzing cleaning and seepage.

For presentation at the EPA/EPRI symposium, "Particulate Control in Energy Processes," San Francisco, 11-13 May 1976

FABRIC FILTRATION PERFORMANCE MODEL

Introduction

This presentation indicates our progress toward development of a predictive model for industrial fabric filtration, with special emphasis on the use of fabric filtration to control the particulate emissions of coal-fired power plants. Primary goals are the prediction of collection efficiency versus particle size and the relationship between the flow per unit area and the pressure drop for a fabric filtration installation.

The following factors can be expected to influence efficiency and the pressure/flow relationship:

1. Dust — chemical composition, particle size distribution, particle shape, particle phase, particle concentration.
2. Gas — chemical composition (especially moisture content), temperature, pressure.
3. Fabric — material, weave (including fill and warp counts), finish, history (especially accumulated dust).
4. Cleaning operations — cleaning type, intensity, frequency, duration.

Ideally, we would like to be able to predict fabric filter drag as well as efficiency for all combinations of the above, but this is a very ambitious task. In forming a useful model, one treads a narrow path between untractable complexity and impractical simplicity — this seems particularly difficult in modeling fabric filtration.

Description of a Filtration System. Figure 1 shows a very simple schematic of a fabric filter installation. A dust-laden flow of gas enters the filter installation with a volumetric flow rate Q and a concentration c. The flow is divided among n compartments, the bags within these compartments having average dust weight per unit area of fabric given by W_i. The pressure drop across the i^{th} compartment's bags is given by the equation

$$\Delta p_i = Q_i S_i / A_i, \tag{1}$$

in which Q_i is the volume flow rate, A_i is the cloth face area and S_i is the drag caused by the fabric and any accumulated dust. The ratio S_i/A_i is analogous to electrical resistance (with Q_i analogous to current and Δp_i analogous to voltage). The total flow, Q, is the sum of the individual flows, Q_i (as long as temperature and pressure corrections are made). Usually the compartments are in parallel and so constructed that the pressure drop at any given time is the same across all of them, so that Δp_i just becomes Δp.

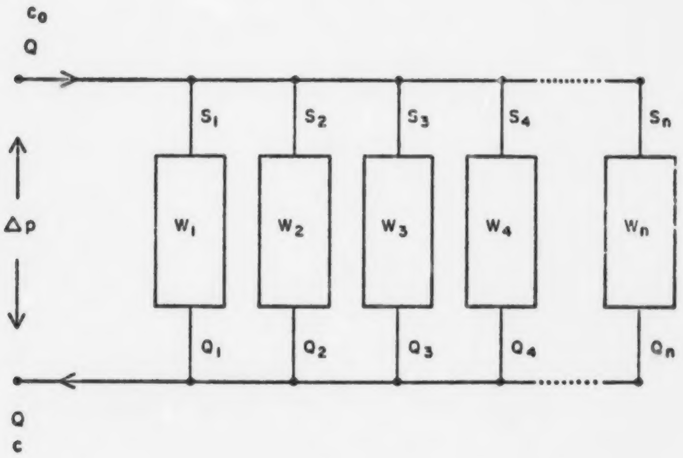

Figure 1. Schematic of n-compartment baghouse.

The relationship between the pressure drop and the volume flow for a particular installation will depend upon the locus of the intersections of the system fan curve and the system resistance curve,[1] each of which can be expressed as volume rate of flow versus pressure drop. Usually, one of the following conditions holds approximately for the installation:

1. The fan produces a constant volume of flow, and pressure drop changes with system resistance.
2. The fan produces a constant pressure drop, and the volume flow changes with system resistance.

Objectives. The cost of the installation will depend upon the fabric chosen and its total area, among other things, and one of the operating costs will be the power consumption due to the pressure drop across the system. Often the available choice of fabrics will be limited, so that the major question becomes the area of cloth needed to handle a volume flow rate of gas.

The cloth area required has as its criterion "operation at an acceptable pressure drop across the cloth for a predetermined cycle."[2] One might add that this assumes that the collection efficiency is adequate under these conditions. Thus one important goal is to be able to predict the pressure drop for a particular dust and fabric combination at a given ratio of air flow to cloth area, the face velocity. The drag will depend upon how much dust is on the filter surface, how it is distributed, its geometrical nature in cake form, the geometry of the fabric, and the viscosity of the gas. A second goal is to predict the emissions from such a system.

Outline of Model. A procedure for calculating the pressure/flow relationship and the efficiency of an installation can be developed by subdividing the fabric area into smaller homogeneous sub-units (compartments, bags, or areas on bags) and then doing the following:

1. Calculate the drag (the pressure drop per unit face velocity) for the sub-unit.
2. Determine the flow from the drag and the instantaneous pressure drop.
3. Determine the penetration, the fraction of the particulate concentration reaching the sub-unit which then penetrates to the clean air side.
4. Calculate the emissions rate from the sub-unit (penetration times inlet concentration times volume flow rate).
5. Calculate the new dust loading of the sub-unit.
6. Determine the new pressure drop or the new total flow rate by combining the resistances of the sub-units according to the law for the addition of parallel resistances:

$$Q = \Delta p \frac{A}{S} = \Delta p \sum_{i=1}^{n} A_i/S_i \qquad (2)$$

$$S = \left[\sum_{i=1}^{n} A_i/S_i \right]^{-1} A \qquad (3)$$

To develop a time profile of the performance, this procedure must be done iteratively, with any cleaning taken into account as well.

Developing this kind of model was our goal. With such a model one should be able to predict the collection efficiency and the relationship between flow and pressure drop for fabric installations for reasonable combinations of variables relating to dust, gas, fabric, and cleaning. Another part of the work at GCA has been laboratory investigations by Dennis et al.,[3] the results of which have influenced aspects of this model.

Models for Drag - Background

Much of the work in filtration has been done on systems other than fabric filters. Of the fabric filtration work, most of the predictive modeling has been directed at forecasting the pressure drop given a fabric and dust combination.

Robinson et al.,[4] for example, produced a three-compartment model. For the increase of drag in any one compartment with increased weight per unit area on the filter, they used the linear drag expression:

$$\frac{\Delta p}{(Q/A)} = S = S_E + K_2 W \tag{4}$$

in which the term S_E is the effective drag, the drag on the fabric obtained by extrapolating the linear portion of the actual drag versus weight curve (S versus W) back to W = 0. By solving the equations for the total drag at constant pressure drop for the three-bag system, Robinson et al.[4] fit the flow rate of the i^{th} bag to:

$$Q_i = a_i t^{x_i} \tag{5}$$

where t_i is the time since last cleaning. They used this flow to obtain the loadings on each bag as a function of time. They combined the individual compartment drags to get the total drag for the system, developing a complicated approximate expression that is only appropriate for a three-compartment system, although the technique can clearly be extended to many-compartment systems.

Solbach[5] also used the linear drag versus weight model, Equation (4). For the case for which the pressure drop, Δp, is fixed but the flow is changing, Solbach[5] expressed the increase of drags due to the increase of the dust loading as

$$dS = K_2 \; c \; V \; dt \tag{6}$$

and integrated this to give the velocity of the gas flowing through a single compartment at any time (t = 0 right after cleaning):

$$V = \Delta p / \left[2K_2 c \Delta p t + S_E{}^2 \right]^{1/2} \tag{7}$$

Solbach[5] extended this to a multicompartment system by averaging V to obtain the expression:

$$V_{av} = \left(\left[2K_2 c t_c \Delta p + S_E{}^2 \right]^{1/2} - S_E \right) \Big/ K_2 c t_c \tag{8}$$

This expression is derived assuming constant pressure drop and infinitely many compartments. It is approximately correct (assuming linear drag versus weight) for systems with much more than a few compartments and even for constant flow in such a system rather than for constant pressure drop. (The reason it is nearly correct for a many-compartment system is that the parallel flow arrangement tends to even out the pressure drop changes.) One obtains the required total filter area from the total volume flow rate and the average face velocity

$$A = Q/V_{av} \tag{9}$$

Dennis and Wilder[6] derived an equation for the average drag for a changing pressure drop:

$$S_{av} = (\Delta p)_{av}/V_{av} = S_E + K_2 W_c/2 \tag{10}$$

$$W_c = cV_{av}\, t_c \tag{11}$$

It can be shown that this result and Solbach's agree: in both cases, the average drag is just the drag at half the final cloth loading (W_c).

This is not an exhaustive list of the investigators who worked with linear models, but it is meant to present the linear drag multi-compartment model. We note that the linear models cannot describe the nonlinear part of the performance curves and there remains the problem of obtaining both parameters K_2 and S_E.

To predict K_2, one approach is to use the Kozeny-Carman equation (or variants), as did Williams et al.,[7] a version of which is

$$K_2 = (25/6)\mu(1-\epsilon)S_p^{\,2}/\rho_p \epsilon^3 \tag{12}$$

where K_2 = cake resistivity, sec^{-1}
μ = gas viscosity, poise
ϵ = filter cake porosity (fractional void volume)
S_p = surface-to-volume ratio of particulate material, cm^{-1}
ρ_p = particle density, g/cm^3.

K_2 is quite dependent upon particle size (increasing for smaller particles) and the porosity of the cake deposit. (Unfortunately, K_2 is often measured in such English units as (inches WC/fpm)/(lb/ft^2).)

Fraser and Foley[9] took a somewhat different approach, keeping the K_2 particle size dependence (and integrating over the particle size distribution), while using correction factors for particle shape, gas velocity and fabric-surface interaction, following Billings and Wilder.[10]

The effective drag, S_E, is really just an extrapolated value, artificial in that such a drag does not occur at that ($W = 0$) condition, so it is not surprising that there is no theory for predicting it, only correlations. Often one must measure it for the dust and fabric of immediate interest.

The only nonlinear model of drag versus weight of which we are aware was presented by Davis et al.[11] which we learned about after beginning our rather different approach. Davis et al.[11] fit equations of the form

$$S = a\ W^b \tag{13}$$

to data obtained for the nonlinear part of the drag versus weight curve. They found correlation coefficients of greater than 0.90 for 104 of 106 data sets. Davis et al.[11] evaluated this at a loading of 0.0244 g/cm^2 (0.05 lb/ft^2), where their drag data generally became linear with W, to obtain S_E. They found that both a and b could be correlated with K_2, but an analysis of variance did not show that K_2 was different for different fabrics for a given dust, which is as expected, even though contradicting the conclusion of Draemel.[12] The model for drag presented by Davis et al.[11] is useful because it includes the nonlinear, cake-repair zone. The model we present also has this feature, with the advantage that the parameters seem more amenable to theoretical prediction rather than just empirical correlation.

Nonlinear Drag Model

The linear models of drag versus weight are incorrect for the low dust loading portion of the filtration cycle, during which the flow and the emissions through the bag are the greatest. The new fabric drag is S_o. The conditioned, cleaned fabric drag is S_R at residual loading W_R. Typically, the drag of the filter increases with added weight per area at one rate, K_R, as the conditioned bag collects dust right after cleaning. This rate (which is the slope of S versus W) changes until it reaches a value K_2, after which the rate remains constant (the S versus W curve becomes linear). The linear part of the performance curve starts at loading W_I. An improvement on the linear model would be one that starts with S_R and K_R, approximates the transition to the linear zone, and matches the linear part of the drag curve.

Viewing the filter as an assemblage of pores (rather than cylinders), we assumed that pores which were unclogged ("open") had a different dust resistivity (K_o) than pores which were caked (closed), assumed to have resistivity K_2, as has the cake. We would expect K_o and K_2 to be different because the velocity in the pores is much higher than through the cake but the particles in the "open" pores are not packed together the way they are in the cake. (Depending upon the magnitudes of such velocity gradient and dust packing effects, K_2 may be greater than K_o or vice versa.) By assuming that pores are either caked or open and that the decrease in the population of open pores is proportional to the number of pores still open and to the increase in fabric loading, dW, we obtained the expression:

$$dN = -(1W^*)\ N\ dW \tag{14}$$

where N = number of open pores per face area of filter

W* = fabric loading that is characteristic of pore-caking.

The fraction of pores still open at loading W becomes:

$$N/N_o = e^{-W/W^*} \tag{15}$$

where N_o is the number of pores per area in the fabric when W = 0.

If the caked pores and open pores contribute to the drag in proportion to their resistivities and their numbers, then it can be shown:

$$\frac{dS}{dW} = K_o\ f(W) + K_2\ g(W) \tag{16}$$

where $f(W) = \exp^{(-W/W^*)}$ (the fraction of pores which are open) and

$g(W) = 1 - f(W)$ (the fraction of pores which are caked).

By integrating Equation 16 we obtain:

$$S = S_o + K_2W + (K_o - K_2)\ W^*\left(1 - e^{-W/W^*}\right) \tag{17a}$$

in which S_o is measured new cloth drag; alternately

$$S = S_R + K_2W' + (K_R - K_2)\ W^*\left(1 - e^{-W'/W^*}\right) \tag{17b}$$

where $W' = W - W_R$, the difference between the loading and the residual loading, W_R, at which $S = S_R$.

We hope to be able to predict K_R and W* from dust and fabric characteristics, but thus far we too have had to rely on correlations. To show how well this form of equation can fit data from an actual fabric, we present Figure 2, for a glass fabric filtering mica dust and cleaned by shaking, and Figure 3, for Dacron filtering fly ash and cleaned by a combination of shaking and reverse air. Both are fabrics appropriate for high-temperature operations. The optimal choice of W* produces excellent agreement with the measured values of drag versus weight. The three other such comparisons we made were similarly successful; Tables 1 and 2 give more information about the systems against which the model was compared in our Interim Report to EPA.[16]

To use this model for a variety of filtration systems, we need theoretical equations or predictive correlations for the parameters K_R, S_R, and W* . . . assuming we have the usual information about the fabric (permeability, from which comes S_o) and dust (resistivity, K_2).

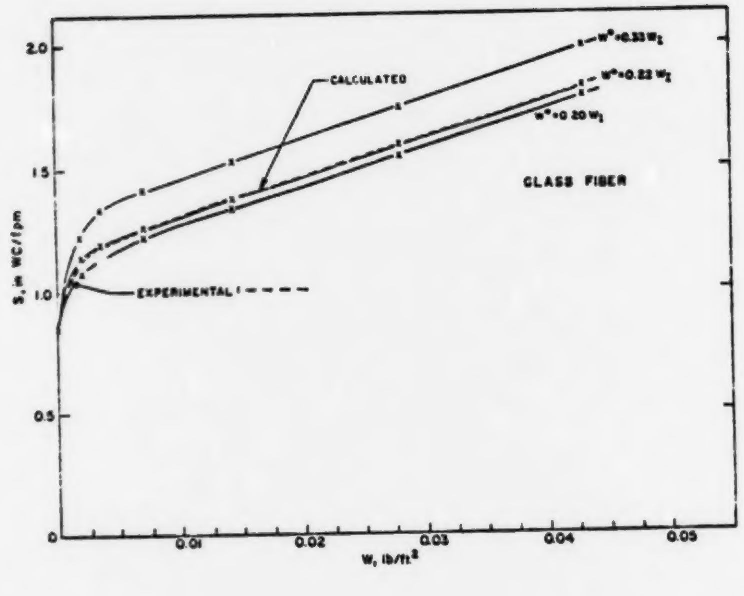

Figure 2. Theoretical and experimental curves of glass fiber filter.

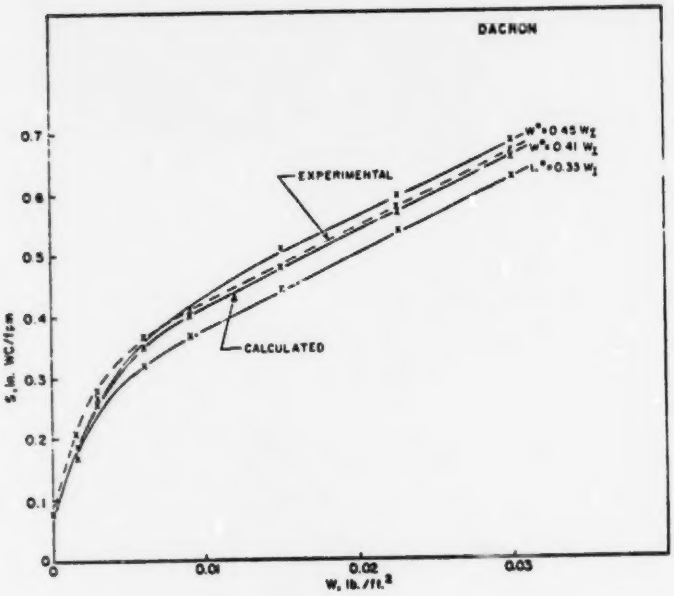

Figure 3. Theoretical and experimental curves of Dacron-type filter.

Table 1. TECHNICAL DATA – FABRIC FILTERS SELECTED FOR VERIFICATION[16]

Type of fabric	Weight, oz/yd^2	Weave and yarn count	Frasier permeability, ft^3/min @ 0.5 "H_2O	Reference
Glass fiber	9.06	3/1 crowfoot, filament 55 x 58	7.9	Spaite and Walsh[13]
Polypropylene	4.30	3 x 1 twill, filament 74 x 33	15.0	Durham[14]
Dacron	10.0	Plain, staple 30 x 28	55.0	Dennis and Wilder[6]
Cotton	10.0	Unnapped sateen 95 x 58	13.0	Dennis and Wilder[6]
Polyacrylester	9.8	2 x 2 twill, spun 39 x 35	60.0	Durham[14]

Table 2. EXPERIMENTAL DATA – FABRIC FILTERS SELECTED FOR VERIFICATION[16]

Fabric type	K_2	K_R	S_R	W_I	Type of cleaning	Dust type	Remarks	Ref.
	" WC-ft-min/lb		" WC/fpm	lb/ft^2				
Glass fiber	16.0	400.0	0.84	0.0036	Shaking	Wet ground mica	Conditioned velocity 2 ft/min	Spaite and Walsh[13]
Polypropylene	6.1	135.4	0.35	0.0135	Shaking	Fly ash	Conditioned velocity 4 ft/min	Durham[14]
Dacron	12.4	93.8	0.08	0.0066	Shaking and reverse air	Fly ash	Used fabric velocity 3 ft/min	Dennis and Wilder[6]
Cotton	15.0	72.0	0.5	0.0075	Shaking and reverse air and shaking only	Fly ash	Velocity 3 ft/min	Dennis and Wilder[6]
Polyacrylester	4.6	26.3	0.05	0.030	Shaking	Fly ash	Conditioned velocity 4 ft/min	Durham[14]

The value of K_R is expected to depend upon the dust, the fabric, and the cleaning, so we thought it might correlate with K_2 for the dust and S_o for the fabric for cleaning by shaking. We calculated (K_2/K_R) for data on glass,[13,15] Dacron,[13,14,6] and Nomex[14] fabrics. (Aerosols of mica and fly ash were used with the glass and the Dacron filters; the Nomex tests were with fly ash.) The correlation was

$$K_R = 515\ K_2\ S_o^{1.035} \tag{18}$$

where S_o is in the units inches WC/fpm and K_R, K_2 are in inches WC-ft-min/lb. Similarly, correlations for the same five fabric-duct combinations gave:

$$S_R = 6.2\ S_o^{\ 1.445}\ K_2^{0.747} \tag{19}$$

(residual drag increases with fabric drag and dust resistivity), and

$$W^* = 7.68 \times 10^{-4}/K_2^{0.913}\ S_o^{0.945} \tag{20}$$

(pore caking requires more material for rather open fabrics and more coarse dusts). Again, the parameters are in English units (W in. lb/ft^2 and K in the units of S/W).

It should be mentioned, at least in passing, that data from the reported measurements of K_2 we used were better correlated with the inverse of particle mass median diameter rather than the inverse squared (the latter is predicted by the Carman-Kozeny equation[16]) and showed a face velocity effect (as has been reported before[17]).

To sum up: the form of our nonlinear expression fit quite well that data for which we evaluated it. The parameters in the model should be theoretically predictable, a problem on which we are working, but until they are, one can use correlations such as those presented here. Such a model promises to be an improvement on the linear model because it can describe that part of the filter performance for which flow and emissions are greatest, the low-loading, nonlinear part.

Efficiency Models - Background

Before presenting our model for woven fabric filtration efficiency as a function of particle size, we will discuss several recent efforts in that direction.

A theory of particle collection by clean cylindrical fibers was adopted by Fraser and Foley[9] for the evaluation of the collection filter efficiency, expressed as a function of three collection mechanisms. They used an equation derived by Davies[18] for low gas velocity ($Re \sim 0.2$) and single cylinders, incorporating impaction, interception and diffusion. Then, they applied a correction factor to the single fiber efficiency γ to account for the effect of the neighboring fibers

$$\gamma_M = (1 + 4.5\,\beta) \tag{21}$$

to obtain the total collection efficiency. The penetration for a fiber was expressed as:

$$Pn = \exp\left[\frac{-4\beta\ \gamma_M\ L}{\pi d_f(1-\beta)}\right] \qquad (22)$$

Due to the difficulties related to the precise measurement of the fiber diameter, they used an effective fiber diameter, derived by Davies:[18]

$$d_f = \left[\frac{70\ \eta_g \beta^{1.5} L(1 + 52\ \beta^{1.5})}{\Delta p}\right]^{1/2} \qquad (23)$$

(The filter packing density is β, its thickness is L, and the gas viscosity is η_g.) That model of the penetration of the particles through the clean fiber filter was extended to the dusted fabric filter under the assumption that the main effect of the dust on the fabric itself is to change the effective fiber diameter. An equation for the change of the thickness of the filtering layer during the filtration cycle was developed by assuming that the dust fills up first the free volume of the filter (at the packing density of the dust) before building up the dust cake. The residual dust loading was expressed as a function of a cleaning parameter and permeability.

The collection efficiency used in that model was based on the calculation of particle trajectories around a single fiber, derived by Davies[18] for a clean fiber with a very small diameter. It is doubtful that the same equation will hold for fiber filters having fiber diameters $\sim 10^2$ μm. If we assume the average face velocity of 2 cm/s (4 ft/min), with fly ash particles having the diameter of 10 μm, fiber diameters approximately 800 μm (which is an average yarn size) and filter packing density around 0.3, then the Stokes number equals 0.02, a lower value than the critical Stokes number, below which the impaction mechanism does not contribute to collection efficiency. (Also under such conditions the diffusion parameter is negligible.) Then the penetration for the particles would be 0.99, considering the interference effect of neighboring fibers to be $(1 + 4.5\ \beta)$. Modeling the fabric filter as an array of cylinders does not seem successful thus far.

Leith and First Model. By using tagged fly ash aerosols, Leith and First[19] were able to distinguish between those fly ash particles which, under laboratory conditions, penetrated a needled felt fabric filter immediately and those particles which exhibited a delay in their penetration. These researchers postulated three types of penetration mechanisms: direct penetration, gradual seepage of the dust, and the breakage and penetration of plugs of material in the vicinity of pinholes.

The experiments were done at face velocities from 5 cm/s (10 fpm) to 15 cm/s (30 fpm) for dust cakes up to 60 μm thick. These velocities are higher than those in normal use for the filtration of fly ash (~ 1 cm/s) and these cake thicknesses are rather low. The efficiency as a function of particle size was such that the penetration was found to "remain relatively constant for particles from 0.3 to 4.0 micrometers in diameter." The penetration increased with face velocity. From their figures for mass flux versus time (at 10 cm/sec) we conclude:

1. Straight through mass flux seemed to decrease (roughly exponentially) with deposit thickness, but may have gone through a minimum near 20 μm thickness.
2. Seepage mass flux remained fairly constant with time and deposit thickness.
3. Pinhole plug mass flux decreased with increasing deposit thickness.

The fraction of the total penetration which was due to pinhole plugs and seepage was greater than the direct (straight through) contribution (at 60 μm thickness) for 15 cm/s but substantially less than the direct for 10 cm/s, which suggests the emissions due to seepage and pinhole plugs might be very much less than the direct at 1 cm/s (2 fpm).

Leith and First[19] postulated the following functional forms for the three types of penetration:

1. Direct penetration proportional to e^{-aX^b}, where X is the deposit thickness
2. Constant seepage mass flux versus thickness of deposit
3. Pinhole plug mass flux proportional to Xe^{-cX} (in which a,b,c, are proportionality factors).

The correlations they found in using these equations ranged from 0.86 to 0.91 correlation coefficients, for mechanism-by-mechanism comparison, adding support for their proposed mechanisms.

Leith and First[19] found only a weak particle size dependence for efficiency, with the efficiency decreasing slightly as particle size increased. The particle size dependence, the velocity dependence, and their general appraisal of the filtration process led them to conclude:

> "Because media filtration theory does not describe the trends in penetration found in a fabric filter, and was not developed for the operational conditions found there, it should not be used to predict or interpret the penetration characteristics of fabric filters."

This conclusion underscored the need for a new approach to the prediction of fabric filter efficiencies, especially for woven fabrics.

A New Model for Fabric Filter Collection Efficiency

As indicated by our review of former models, most treatments of filter collection efficiency were not for fabric filters but for other kinds of filters. The synthesis attempted for fabric filters by Fraser and Foley[9] used as its basis formulas for collection on cylinders. Because of the overlapping caused by weaving and the closeness of the yarns, the flow in a fabric filter is very unlike that of a cylinder in a uniform flow stream, the starting point for models of filtration based on collection by cylinders. Rather, the fabric produces a flow which is directed through the interstices of the weave at and near the intersections of the yarns. This much more resembles flow to an array of apertures than it does flow around single or grouped cylinders. Because the nature of the flow has a strong bearing on the collection efficiency, one expects a pore model to be inherently better than models which start from a cylinder or an array of cylinders.

Our analysis fits into a baghouse simulation model outlined in the flow chart in Figure 4. Once the starting data are put into the simulation, the program calculates flow, pressure drop, fabric loading, penetration, and emissions as a function of time for the bags in the simulated baghouse. This section discusses the equations and analysis which went into calculating the emissions.

Collection Mechanisms. The model can calculate the fraction of incoming particles which are captured in the cake, in the fabric, and in the pinholes which go through cake and fabric. The collection mechanisms modelled are impaction, interception, and diffusion.

Collection by Impaction, Interception, and Diffusion in the Cake. The particulate material collected by the fabric filter builds up first in the vicinity of the pores. As it builds, this deposit starts to form a cake through which a substantial portion of the flow passes, and particles are captured by the particulate matter of the cake. We seek to model the capture of particles by the filter cake. The geometry is quite complex where polydisperse aerosols are involved. The uncertainties are so great with regard to geometry that the available information warrants only an approximate model. We have formed such a model on the basis of analysis of characteristic time scales, characteristic lengths, etc.

An idealized view of the filter cake is shown in Figure 5. The cake is viewed as a series of layers, with a total thickness L, in cm. The cake thickness is given by

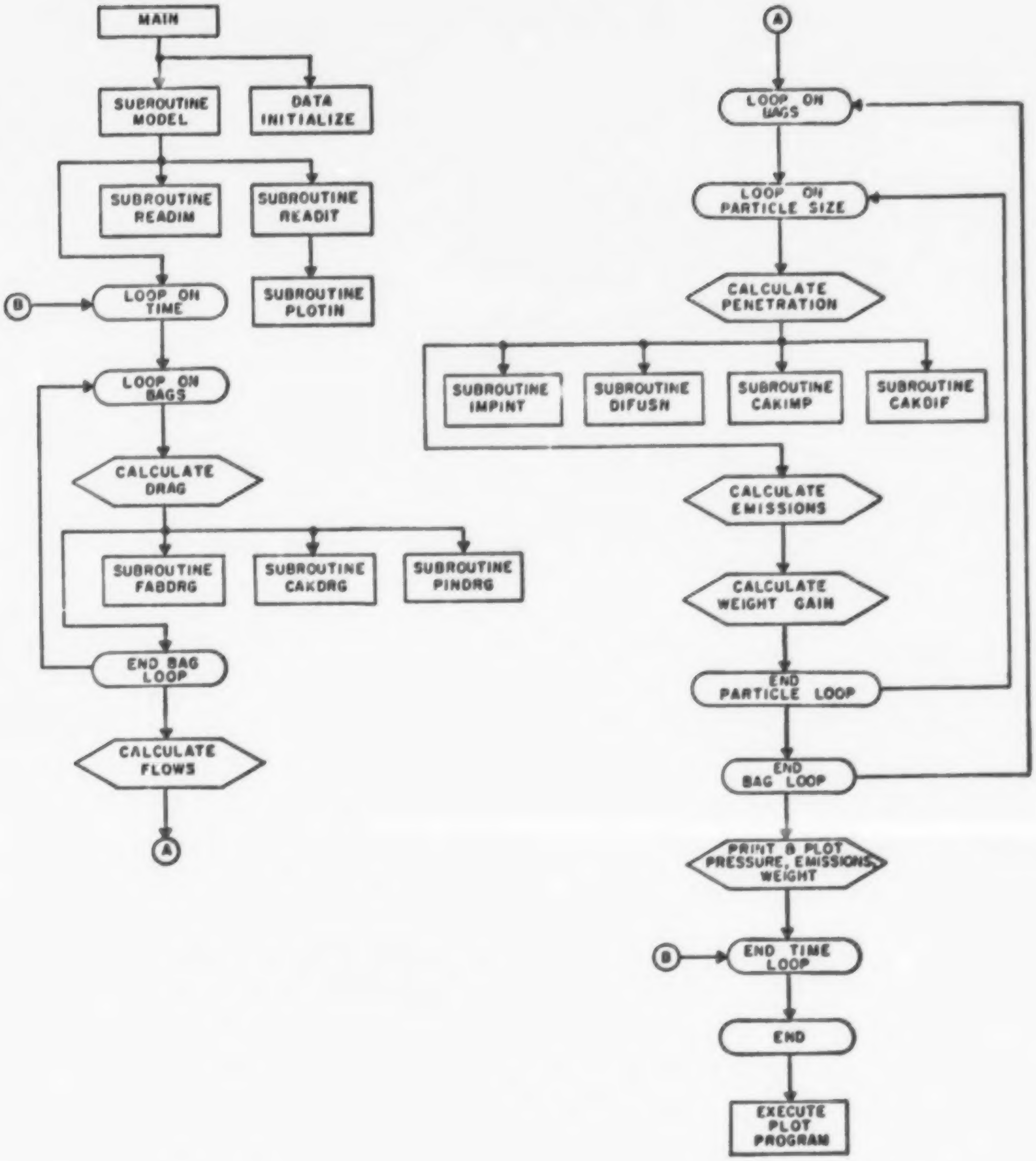

Figure 4. Baghouse simulation program flow chart.

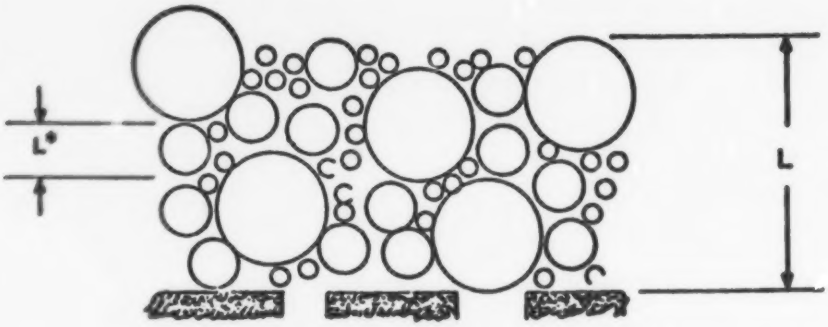

Figure 5. Fabric filter cake, resting on fabric.

$$L = W/\rho_p(1 - \varepsilon) \tag{24}$$

in which W = particle loading, g/cm^2

ρ_p = particle material density, g/cm^3

ε = void volume fraction, porosity.

By taking the number of particles and dividing it into the volume of the cake, we can get the mean volume of cake per particle, thus the mean thickness of each layer:

$$L^* = \left[\pi/6(1 - \varepsilon)\right]^{1/3} \text{VMD} \tag{25}$$

in which VMD = particle volume mean diameter, cm.

Assuming that the number of void volumes in the cake is equal to the number of particles, we have the following void volume mean diameter, d_v:

$$d_v = \left[\varepsilon/(1 - \varepsilon)\right]^{1/3} \text{VMD} \tag{26}$$

The mean velocity in the open portions of the cake is given by:

$$\overline{U} = v_f/\varepsilon \tag{27}$$

in which v_f = face velocity, cm/s.

A characteristic velocity for flow through the cake is the face velocity divided by the porosity, $\overline{U}$. A characteristic dimension over which the flow changes direction is the mean diameter of a void, d_v.

The inertial parameter (Stk, the Stokes number) can be defined from these equations as:

$$\text{Stk} = \overline{U}\tau/d_v \tag{28}$$

This parameter determines how large inertial impaction may be. We make the simple assumption (e.g., see Fuchs[20]) that the impaction efficiency of a single cake layer is approximately equal to the Stokes parameter at each void volume (up to Stk = 1):

$$E_{Ii} \doteq \text{Stk} \; (E_{Ii} \leq 1) \qquad (29)$$

The likelihood that a particle will penetrate through one void volume is just one minus the efficiency:

$$Pn_{Ii} = 1 - \bar{U}\tau/d_v \qquad (30)$$

and the likelihood of penetrating M such volumes (i.e., through M layers) is:

$$Pn_I = (1 - Pn_{Ii})^M. \qquad (31)$$

(This assumes that the collection efficiency of each void volume in the cake through which the aerosol flows is independent of the earlier portions of the flow path.) The number of such layers is approximately the thickness of the cake, L, divided by the mean distance between particles forming the cake, L*, the mean layer thickness:

$$M \doteq L/L^*. \qquad (32)$$

Interception is the collection contribution due to solely the size of the particles compared with the dimensions of the flow channels. Even if the particles follow the streamlines of the flow, those with centers on streamlines which come within a particle radius of an obstacle will strike the obstacle and be captured (it is assumed). We can approximate the likelihood of penetration in passing through the ith layer as that for passage through a cylindrical channel of diameter d_v:

$$Pn_{Ci} = (1 - d_p/d_v)^2 \qquad (33)$$

and from this we derive an approximate expression for the penetration of a particle through the dust cake of thickness L by:

$$Pn_C = (1 - d_p/d_v)^{2\,L/L^*}. \qquad (34)$$

The Brownian motion of particles causes them to diffuse in a concentration gradient so as to transport particles from higher concentration regions to lower. In the presence of a surface to which the particles adhere, there will be a net diffusive flux of particles to the surface due to this mechanism. The equation governing this diffusion for a flowing aerosol is given by:[8]

$$\frac{\partial n}{\partial t} + \vec{v} \cdot \vec{\nabla} n = D\, \vec{\nabla}^2 n \qquad (35)$$

in which n = number concentration of particles of diameter d_p

$\vec{v}$ = flow velocity

D = particle diffusivity

$\vec{\nabla} n$ = gradient operator

$\vec{\nabla}^2$ = Laplacian operator

For systems of interest it is usually true that the convective term, $\vec{v} \cdot \Delta\vec{n}$, is negligible. Within a void volume this equation becomes approximately:

$$n \doteq D\, t/(0.5\, d_v)^2 \tag{36}$$

for the capture due to diffusion. The time, Δt, is the characteristic flow length divided by a characteristic flow:

$$\Delta t = d_v/\bar{U} \tag{37}$$

As before, for M (=L/L*) layers in the cake, the penetration becomes:

$$Pn_D = (1 - 4\, D/U\, d_v)^{L/L^*}. \tag{38}$$

As Fuchs has noted,[20] there is no simple way for combining correctly the collection efficiency terms (or, equivalently, the penetration expressions). Commonly, the mechanisms are treated as operating independently, in which case the penetration becomes the product of the penetrations:

$$Pn = Pn_I\, Pn_C\, Pn_D\,. \tag{39}$$

We have used this approach in our model.

Impaction and Interception at a Pore or Pinhole. Near the intersections of the fabric yarns are openings which we call pores, the geometries for which are discussed in the Interim Report by Dennis et al.[3] and, to an extent, in Appendix A of our report.[16] It has been observed that often a fraction of such pores, and/or pores caused by other means, stay open during the filtration cycle. We denote them "pinholes." Our model simplifies both pores and pinholes into circular cylindrical channels in the fabric (and in the cake as well, for pinholes). The task then becomes to calculate the collection efficiency of such an array of channels.

Spurny et al.[21] presented calculations for a somewhat similar situation in their analysis of membrane and nuclepore filters, modeled as a highly porous array of parallel circular cylindrical channels. It can be shown that their equations for impaction do not go to the correct limits for low-porosity structures, which is our situation for fabric filter having a few percent or less free area.

Smith and Phillips[22] found numerical solutions to the problem of the collection efficiency of an aperture due to impaction and interception, but they included only capture at the face of the material having the orifice and not along the walls of the channel, which can be shown to be substantial in our context.

Thus, although we could use equations available in the literature for collection due to diffusion in a channel, we did not have applicable equations for impaction and interception of particles by the pores and pinholes and had to derive approximate expressions, discussed next.

Figure 6 shows the geometry of the model. The pore (channel, orifice, etc.) has a radius (or hydraulic radius) R^* and a length L_p. The opening is one of several and the area πR_o^2 is the total face area divided by the number of pores, distributed approximately uniformly. The face velocity, v_f, far away from the pore is just the flow through the pore, Q_p, divided by the area per pore:

$$v_f = Q_p/\pi R_o^2 \qquad (40)$$

The flow is simplified by dividing it into three regions (see Figure 6):

I. In this region (radial distance, r, greater than R_o) there is a uniform face velocity toward the plane of the opening, and the face velocity is given by:

$$v_f = Q_p/\pi R_o^2 \qquad (41)$$

II. In this region (radial distances: $R^* \leq r < R_o$) the flow is the same as it would be in potential flow to a point sink, being directed radially toward the center of the pore face, with the velocity at distance r governed by:

$$v(r) = v^*(R^{*2}/r^2) \qquad (42)$$

in which

$$v^* = Q_p/2\pi R^{*2} \qquad (43)$$

III. In the third region ($r < R^*$) and within the channel there is assumed a uniform flow profile.

Of course, this set of flow profile assumptions is somewhat different from what actually would take place. Still, far away from the pore opening the flow would generally be uniform; there would be an intermediate range for which the flow

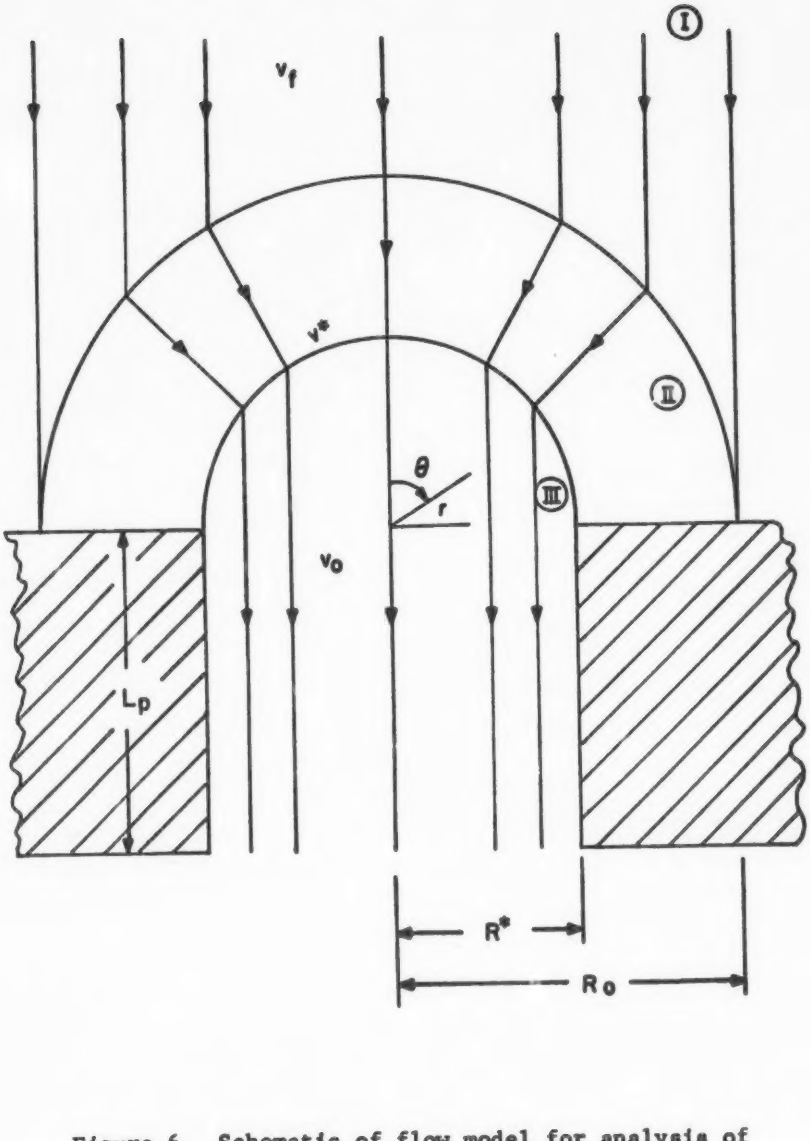

Figure 6. Schematic of flow model for analysis of impaction and interception.

would be approximately that toward a point sink at the center of the orifice; and converging flow at the orifice would be expected to produce a rather uniform flow profile near the entrance before the typical parabolic flow was established for laminar flow conditions. Thus the flow model preserves many of the features of a more rigorous solution of the flow velocity field.

A particle with an initial velocity of v_f will travel a distance L'_{sI} in a stationary medium, with L'_{sI} given by:[20]

$$L'_{sI} = L_{sI}(1 - e^{-t/\tau}) \tag{44}$$

where

$$L_{sI} = \tau \, v_f \tag{45}$$

$$\tau = C\rho_p \, d_p^{\,2}/18\mu \tag{46}$$

in which C is the Cunningham slip correction, ρ_p is the particle density, and d_p is the particle diameter. The variables L_s and τ are the particle stopping distance and particle relaxation time,[20] and t is the time.

We will not give more details of the rather complex method used by us to obtain efficiency due to impaction. Essentially it involved comparing the stopping distances (L_s', including mean residence times in regions II and III) with the distance a particle would have to travel to strike the face of the pore or the inner wall, keeping track of particles which were captured in one region so as not to double-count them in capture in subsequent regions.

The other mechanism considered was diffusion to pore and pinhole walls. The formulas for the decrease in particle concentration in traversing a tube while diffusing have been presented by Fuchs.[20] The concentration decreases approximately exponentially with the parameter μ as the argument:

$$\mu = D \, L_p/V_o \, R^{*2} \tag{47}$$

in which the particle diffusivity is D. The approximate formula for μ much less than 1 is:[20]

$$Pn = 1 - E = 1 - 2.56 \, \mu^{2/3} + 1.2 \, \mu + 0.177 \, \mu^{4/3} \tag{48}$$

Summary of Model Equations. The essential aspects are the collection mechanisms of impaction, interception, and diffusion as applied to the filter fabric, the cake, and pinholes in fabric and/or cake.

Results of Modeling of Fabric Filtration Systems

Introduction. This portion will present some of the results we have obtained in modeling fabric filtration systems with an initial emphasis on coal-fired power plants using fabric filtration, such as the Sunbury plant that was one of the subjects of the Interim Report by Dennis et al.[3] The order is as follows:

1. Q and Δp versus time for 15-bag system, using a linear drag model.
2. Collection efficiency versus particle size and time for a single bag with linear drag.
3. Collection efficiency versus particle size and time for a single bag with nonlinear drag.
4. Collection efficiency versus particle size and time for a three-bag system, using nonlinear drag.

Flow Modeling. In reference 16 we presented the computer program for our model. The drag equation used was linear:

$$S = S_E + K_2 W \tag{49}$$

$$W = cVT \tag{50}$$

where S_E = 0.5 in. WC/fpm = 2.5 cm WC/(cm/s) = 250(N/m^2)/(cm/s)
K_2 = 11 in. WC/fpm/lb/ft^2
= 1.1 (N/m^2)/(cm/s)(g/m^2)
c = 5 gr/ft^3 = 11.5 g/m^3
t = time since cleaning, minutes.

Either V was constant (2 fpm = 1cm/s face velocity) or it was obtained from the pressure drop and the drag. Each bag was cleaned instantaneously (to the residual weight level) every 30 minutes, one bag cleaned per 2 minutes. The results are shown in Figures 7 and 8, where the information is plotted every minute so that we are getting an instantaneous reading at two points in the cycle, just after cleaning and midway between cleanings. From Figure 7, we see that after about two complete cycles (60 minutes), the pressure drop has come into dynamic equilibrium, with a minimum near 1.4, a midpoint about 1.45, and an (inferred) maximum about 1.5 in. WC (375 N/m^2). For 15 bags, the range of pressure drop values is much less than for one bag, as we expected.

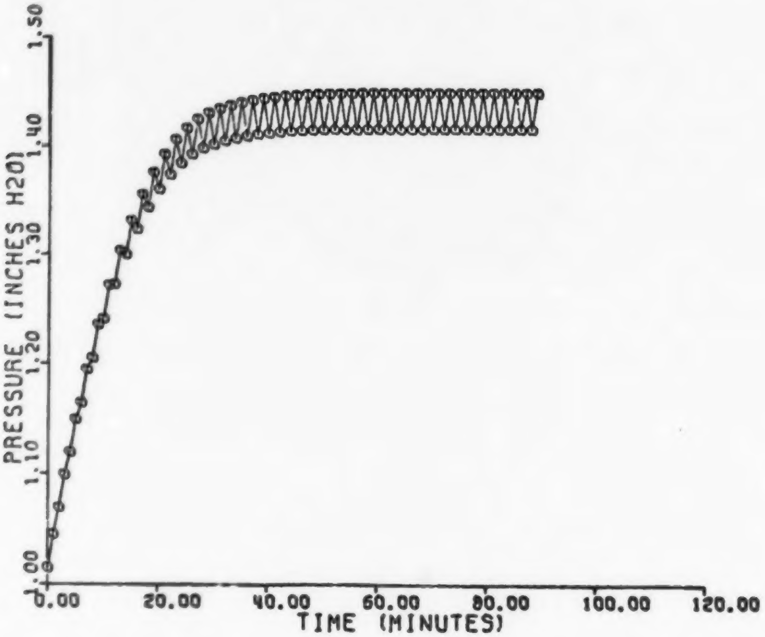

Figure 7. Test run No. 5: 15-bag simulation pressure versus time graph.

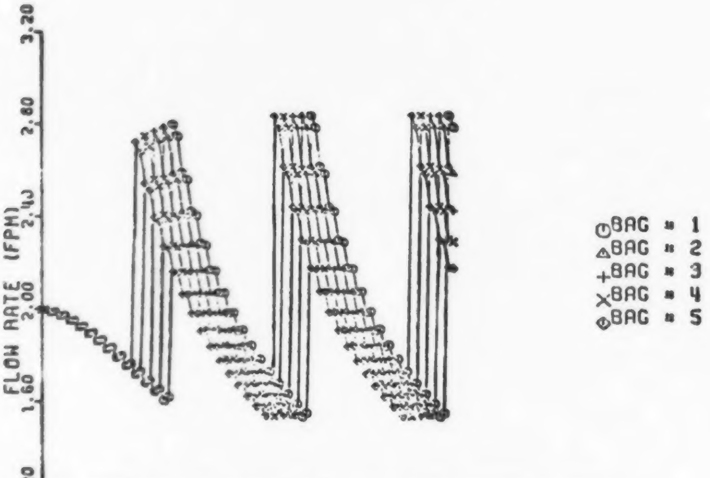

Figure 8. Test run No. 5: baghouse simulation individual flow rate graph.

Because the many-bag system greatly reduces the pressure drop variations with time, we can approximate this new system as follows: even though the total system is one with a constant total flow, it behaves, bag-by-bag, almost as though each bag is kept at a constant pressure drop (and has a variable flow). This is significant, in turn, because we have available to us the general expression governing the relationship between drag and time (thus flow and time) for a bag with constant pressure drop and variable flow, which we presented as Equation 8, from Solbach.^{5}

Figure 8 shows the flow rate for bags No. 5 to No. 1 (the cleaning order was from No. 15 to No. 1). The total flow rate remained constant, but the individual flow rates declined for these bags as they accumulated more weight than the other bags in the system, which were being cleaned before them. When it became the turn of bag No. 5 to be cleaned, its flow rate increased, while the other continued to drop. This process repeated itself with the other bags, and it is interesting to note that their flow rate versus time curves all look very much the same except for being displaced in time.

Efficiency Modeling. Figure 9 shows the results of modeling the collection efficiency of a single bag with the program discussed above and shown in Figure 4.

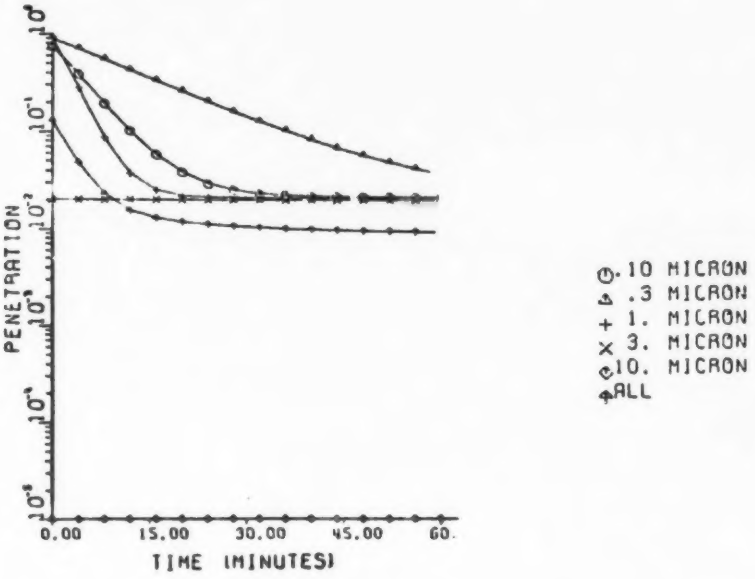

Figure 9. Test run No. 13a baghouse simulation penetration.

The conditions for these results are listed as part of the program output, shown in Table 3. Volume flow rate per area was 2 fpm (1 cm/s). The pore (fabric opening) sphere of influence radius is just the radius of a circle of area equal to the area per pore in the fabric. The pore radius is the approximate hydraulic radius of the pores. The pore length is the yarn thickness. Density is the particle density ρ_p,

Table 3. TEST RUN NO. 13A BAGHOUSE SIMULATION INPUT DATA

Number of bags:	1 bag
Cycle time:	60 min.
Number of increments per cycle:	15 increments
Velocity, Q/A:	1 cm/s
Concentration:	1.142×10^{-5} g/cm^3
Initial drag resistance, S_E:	2.5×10^3 dyn/cm^2/cm/s
Cake resistance coefficient, K:	1.12×10^5 dyn/cm^2/cm/s/g/cm^2
Pore sphere of influence radius:	4.78×10^{-2} cm
Pore radius:	2.5×10^{-3} cm
Pore length:	4.0×10^{-2} cm
Yarn radius:	2.0×10^{-2} cm
Particle density:	1.0 g/cm^3
Gas viscosity:	2.08×10^{-4} poise
Cake porosity:	0.5
Temperature:	350°K
Caked area (uncleaned fraction):	0.0
Pinhole radius of influence:	4.78×10^{-2} cm
Pinhole radius:	5.0×10^{-3} cm
Particle volume mean diameter:	5.0×10^{-4} cm
Number of pinholes per cm^2:	0.31

Particle information:

Diameter (cm)	Distribution	Slip correction	Diffusion constant	Tau
1.0×10^{-5}	0.00	2.87	7.02×10^{-6}	7.79×10^{-8}
3.0×10^{-5}	0.01	1.56	1.27×10^{-6}	3.82×10^{-7}
1.0×10^{-4}	0.12	1.16	2.85×10^{-7}	3.16×10^{-6}
3.0×10^{-4}	0.32	1.05	8.51×10^{-8}	2.58×10^{-5}
1.0×10^{-3}	0.38	1.02	2.49×10^{-8}	2.75×10^{-4}
3.0×10^{-3}	0.17	1.01	8.21×10^{-9}	2.46×10^{-3}

chosen to be one (the density of fly ash would have been a better choice). S_o and K_o were not used, but rather S_E and K_2 were the same values as above. Gas viscosity was that appropriate for temperature of 350°K (= 77°C = 171°F); a higher temperature would probably be more appropriate for coal-fired systems. The caked area (fraction of filter from which the cake is not removed by cleaning) was set to zero. The pinholes, which never close during the filtration cycle, were given the same radius of influence as the pores, with a radius estimated as twice the size of a typical pore, and with a number per area corresponding to about 2 per square inch (see Dennis et al.[3]) The cake dust was assumed to have a volume mean diameter of 5 μm and a porosity of 0.5. In Table 4 we give the approximate size distribution for the Sunbury fly ash (Dennis et al.[3]) used to approximate the inlet particle size distribution; this is log normal with a mass median aerodynamic diameter of 6μm and a geometric standard deviation of 3.0. The values 0.1, 0.3, 1.0, 3.0, 10, 30 μm have been used as the midpoints of equal logarithmic intervals (geometric means of the intervals) for the size distribution. The penetration results in Figure 9 are plotted on semilogarithmic scale. The penetrations are weighted by the mass fraction values in Table 4 to give the curve marked "all particles."

Table 4. APPROXIMATE SUNBURY FLY ASH SIZE DISTRIBUTION (log normal, $d_{ae50} = 6$, $\sigma_g = 3$)

Size interval (μm)	Logarithmic midpoint, geometric mean (μm)	Aerosol mass in size interval (%)
17 -52	30	17
5.2 -17	10	38
1.7 -5.2	3.0	32
0.52 -1.7	1.0	12
0.17 -0.52	0.3	1
0.052-0.17	0.1	<< 1

The particle size with the greatest penetration (of those tested) is 0.3 μm diameter, which is in accord with experimental results from many different aerosol filtration studies. The initially linear slope on this scale (semilogarithmic) is consistent with an exponential decay as the cake gets thicker (recall this is constant velocity, thus the rate of cake build-up is constant). Once the region is reached where much of the flow and virtually all of the penetration is that through the pinholes which do not cake, the penetration becomes nearly constant with cake thickness. A substantial fraction of the 0.3 and 1.0 μm particles penetrate the clean fabric, but then each of these decreases toward an asymptote as the pinhole penetration predominates. The 3 μm particles apparently penetrate only the pinholes and this penetration does not change appreciably during the cycle. The 10 and 30 μm particles do not even penetrate the fabric or the pinholes at all ($Pn < 10^{-5}$) so that the cake buildup does not affect them; they are caught by inertial impaction and by interception by fabric pores and by pinholes. The mass penetration, "all particles," starts at about 10 percent and rapidly declines to about 1 percent. The leveling off of the curve for total mass penetration indicates that this penetration is almost wholly due to pinholes, and it will thus be quite sensitive to the assumptions made about the pinhole population. (The pressure drop versus time, which follows the linear model, went from 2.5 cm WC (1 in. WC) to about 7 cm WC (about 2.7 in. WC), reasonable values.)

Figure 10 has penetration versus time curves for the same situation except that the nonlinear model of drag was used. We started with an initial drag which was roughly that of the Sunbury fabric (S_0 = 0.05 in. WC/fpm), then used the correlations developed above to get the ratio of K_2 to K_R from S_o, K_R/K_2 = 12.5, and to get W* = 0.004 lb/ft^2 = 0.002 g/cm^2. The efficiency curves are very similar to those for the linear drag model, except that the initial rapid change in pressure drop somewhat changes the distribution of the flow between the fabric plus cake and the pinholes, shifting the flow more rapidly to the pinholes, thus producing a bit of increase in the penetration of the 3 μm particles initially, before their penetration becomes effectively independent of time. Again, major features include the maximum penetration by the 0.3 μm diameter particles, the initial exponential decrease for those particles being filtered out primarily by the cake, and the important role of the pinholes as the cake becomes thick.

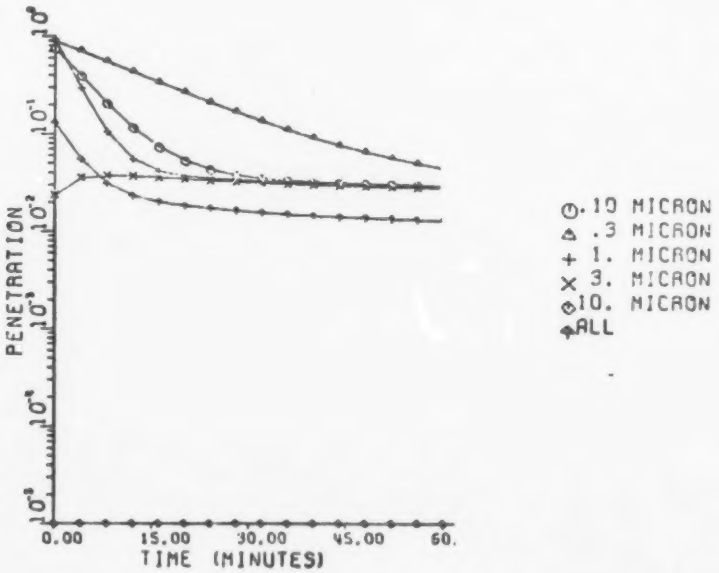

Figure 10. Test run No. 13b: baghouse simulation penetration versus time graph; single bag, nonlinear drag, with pinholes.

To demonstrate further the impact of the pinholes, we repeated the last test (nonlinear drag, etc.) but removed the pinholes, so that all the flow is through fabric pores or cake only. Figure 11 is the penetration without pinholes. The most marked difference is that collection efficiency for all particle sizes now continues to decrease as the cake thickens, and this efficiency becomes better than 99 percent by mass for the particles within 10 minutes. The 0.3 μm penetration is still greatest. The 0.1, 0.3, and 1.0 μm particles all show an exponential decrease with cake thickness now. The 3 μm particles show more rapid changes in collection at the beginning part of the cycle, where pressure drop is changing most rapidly, and this is typical for other collector where inertial impaction is the predominant mechanism, as it is for these 3 μm particles.

The final figure in this series is Figure 12. Here we have used three bags rather than one, the same nonlinear drag parameters as before, and we have restored the pinholes. First, the bags all have the same flow rate, then after one is cleaned, only two have the same flow rate, and the third is much higher in flow. Finally, all three have been cleaned once, and a pattern develops which will repeat itself.

Figure 12 shows the changing pattern for particle penetration versus time with particle size as a parameter. The relative order of penetration with respect to size is the same, and we also still see the change of the penetration for the 3 μm particles as pressure drop changes. Finally, the asymptote provided by the pinhole mechanism is evident. The three-bag system with nonlinear drag produces nearly the same total mass penetration as the one-bag system after 15 minutes, but the cleaning process then returns one of the bags to low efficiency and high flow rate, producing a major increase in penetration for the total system.

Calculator Model for Drag. In some cases, a quick and approximate estimation of an average air-to-cloth ratio of a baghouse unit is necessary or satisfactory. Therefore, we tried to derive a simple equation for calculating an average air-to-cloth ratio. For our derivation we used an expression for the average face velocity (air-to-cloth ratio), $\overline{V}$, which was derived by Solbach[5] and independently also by Dennis and Wilder.[6] This expression is based on the linear drag model. The average face velocity is given by Equation (8):

$$\overline{V} = \frac{1}{K_2 \, ct_c} \left[\sqrt{2K_2 \, ct_c \, \Delta p + S_E^2} - S_E \right] \qquad (51)$$

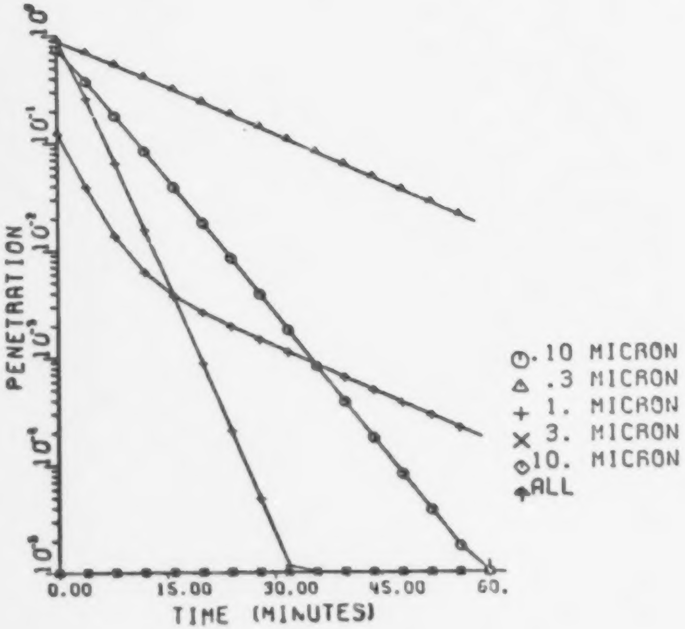

Figure 11. Test run No. 15 baghouse simulation penetration versus time graph; single bag, nonlinear drag, without pinholes.

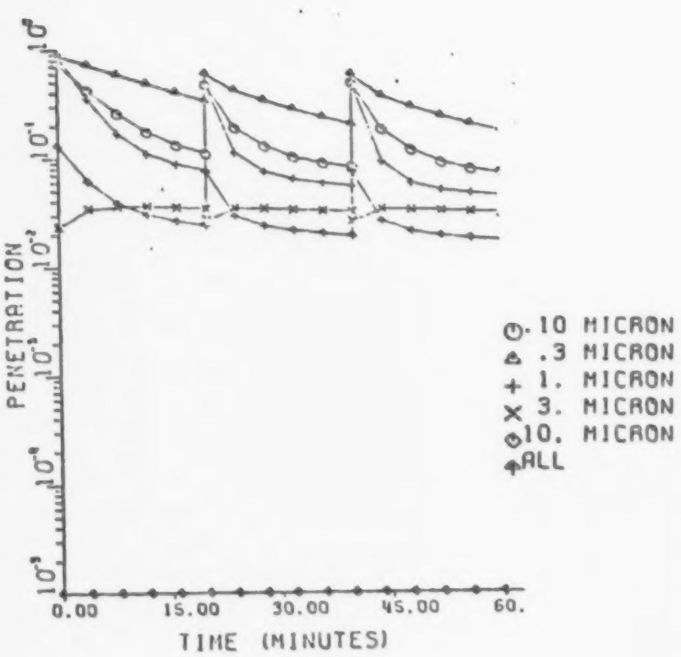

Figure 12. Test run No. 16 baghouse simulation penetration versus time graph; three bags, nonlinear drag, with pinholes.

where K_2 = dust resistivity, in. WC-min-ft/lb
c = dust concentration, lb/ft^3
t_c = total filtration time, min
Δp = pressure drop across the baghouse, in. WC
S_E = effective drag, in. WC/fpm.

If we express the effective drag, S_E, as a function of the clean fabric drag, S_o, by:[16]

$$S_E = 7.4\ S_o^{\ 0.672}, \tag{52}$$

and K_2 as a function of particle mass median diameter, d_p, by:[16]

$$K_2 = 142.4\ d_p^{\ -1.06}, \tag{53}$$

the Equation (51) becomes:

$$\overline{V} = \frac{-\ 7.4\ S_o^{\ 0.672} + \sqrt{54.76\ S_o^{\ 1.344} + 284.8\ d_p^{\ -1.06}\ \Delta p\ ct_c}}{142.4\ d_p^{\ -1.06}\ ct_c}\ . \tag{54}$$

Table 5. PARAMETERS FOR FIVE BOILERS USING FABRIC FILTRATION

Plant	Clean cloth Frazier permeability, fpm @ 0.5" H2O	Dust concentration, gr/ft^3	Total filtration cycle time, min	Pressure drop, " H2O	Particle mass median diameter, μm	Operating velocity, fpm	Predicted velocity, $\underline{V}$, fpm
Pennsylvania Power and Light Co., Sunbury Station, Pa.	54.3	2.	33	2.5	8	1.92	3.90
Colorado UTE Electric Assoc., Nucla Power Station, Colo.	86.2	1.93	29.4	4.2	8	2.79	6.50
Sorg Paper Co., Middletown, Ohio	55.0	3.2	30	4 - 5	12	2.07	6.10
Crisp County Power Commission, Cordele, Georgia	55.	2.6	60	4.	12	2.2	4.69
Pennsylvania Power and Light Co., Holtwood Station, Pa.	45 - 65 (50)	7.7	33	6.5	10[a]	2.26	4.60

[a] Assumed values.

Then, using industrial data for Δp, d_p, S_o, c and t_c, shown in Table 5, we calculated the face velocity (air-to-cloth ratio) for each baghouse. These values are shown in Table 5 along with the operating industrial values of the face velocity. (Sources of the data are listed elsewhere.[16]) The ratios of operating to predicted velocities are very close. The average of all values of that ratio was 2.29 so that we have a corrected estimate, $\overline{V}'$, of the average face velocity:

$$\overline{V}' = \overline{V}/2.29 \tag{55}$$

The differences between corrected predicted and operating values were within ± 15 percent. More data would be needed to improve the reliability of our last equation, obviously.

Final Comments

What is presented here seems the most detailed model yet of a fabric filtration system, and it has sufficient flexibility to have added to it a number of refinements as our information warrants.

The areas which seem most in need of further study are cleaning and seepage. Regarding cleaning, we would lime to know the amount and spatial distribution of the residual dust as a function of cleaning type, intensity, duration, and frequency for several dust/fabric combinations, especially those of applicability to controlling fossil-fuel boiler emissions. Regarding seepage, we would like to know what variables affect the rate at which it occurs and what the relationships are. Of course, the application of fabric filtration does not require such information, but the optimization of fabric filtration does.

Accurate methods for prediction of fabric filter drag and collection efficiency would greatly aid the design of optimal fabric filter systems. By viewing the fabric as an array of apertures rather than an array of cylinders, we have been able to derive expressions describing the nonlinear dependence of filter drag on fabric dust loading and the collection efficiency of the fabric and the dust cake as functions of particle size. Much work remains to be done, especially regarding cleaning of fabrics, but we believe that this new approach can provide a useful framework for future development as well as yielding useful insights concerning the factors which determine drag and collection efficiency. As the model is improved and as more data from operating systems are incorporated, it is hoped we will have a method which will facilitate the application of fabric filtration to the problem of obtaining power economically with minimal environmental impact.

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PERFORMANCE AND ENGINEERING EVALUATION OF THE NUCLA BAGHOUSE

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I. Introduction

A. Objective

The objective was the evaluation of the baghouses at the Nucla Station of the Colorado Ute Electric Association with a field performance test and engineering study. This paper is a summary of the study. For more details, the final report by Ensor et al. [1976] should be consulted.

B. Site Description

The Nucla Station is located five miles southeast of Nucla, Colorado, a small town one hundred miles southeast of Grand Junction. The coal, purchased from Peabody, Inc., is mined about ten miles away and is trucked to the plant. The coal has a heating value of about 12,000 Btu/lb, 14-20 percent ash, 45 percent fixed carbon and 0.5 to 0.7 percent sulfur.

The station has three generators, each with a nameplate rating of 12,650 kw. The three Springfield spreader stoker boilers are rated at 139,000 lb/hr of steam at 825°F and 625 psi. Gas from each of the boilers passes through an economizer, air preheater, baghouse, induced draft fan, and out the stack. The gas streams of each of the three boilers are identical and separate. A diagram of the system is shown in Figure 1.

Each boiler has a Wheelabrator-Frye Size 814, Model 264, Series 8, six Module Dustube Dust Collector. The design parameters are summarized in Table 1. Each Collector has six compartments which are independent and can be cleaned or maintained without interference to the other operating compartments. The bags are suspended from the shaker mechanism by a hook and a spring to maintain the desired tension and are connected to the "cell plate" at the bottom by a snapring.

The compartments are cleaned in sequence when the pressure drop across the baghouse reaches about 4.5 in H_2O. The cycle consists of a 1 minute delay, 15 seconds reverse air, 1 minute delay, 10 second shake, 45 second delay, another 15 seconds of reverse air, and 45 second delay for each compartment. The reverse air is obtained from the discharge side of the baghouse by a 30 hp centrifugal blower. The reverse air flow is about 5500 cfm for an air-to-cloth ratio of 1.1 acfm/sq ft. The frequency of the shakers is 4 cycles per second.

II. Test Methods

A. Size Distribution

Cascade Impactors. The impactor used in this study was designed at Meteorology Research, Inc., (MRI) to facilitate sampling of particulate matter in stacks. The design is based on a simple annular arrangement of jet and collectors reported by Cohen and Montan [1967]. The assembly has been constructed using quick disconnect rings to increase the flexibility of

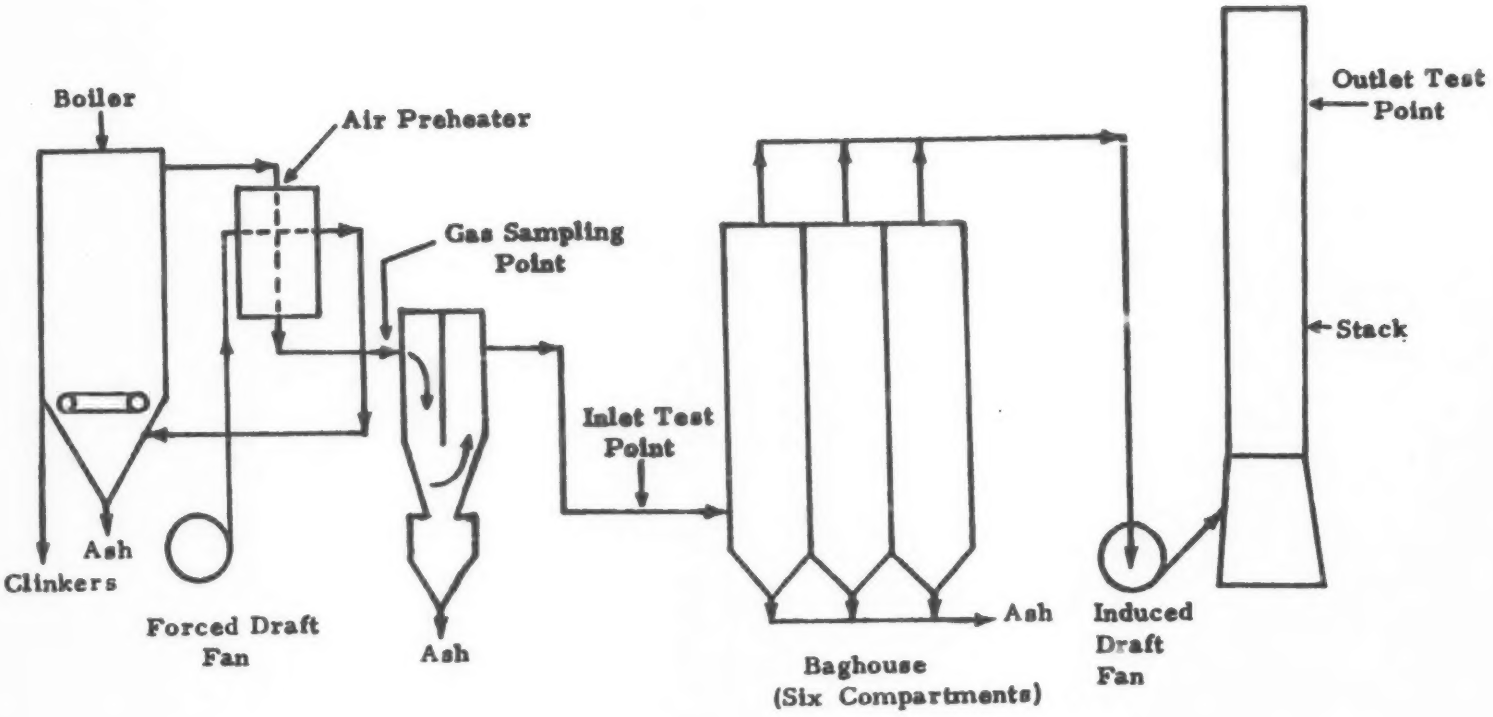

Figure 1. Flow Diagram of Major Gas Streams, one of three identical systems

Table 1. Baghouse Design Parameters

Gas Flow Rate	86,240 acfm
Compartments	6
Bags/Compartment	112
Bag Manufacturer	W. W. Criswell Division Wheelabrator-Frye
Model	No. 445-04 "Dustube"
Diameter	8 in.
Length	22 ft
Active Area/Bag	46.0 ft^2
Fabric	Fiberglass 66 x 30 thread count, 3 x 1 twill
Finish	Silicone/Graphite
Permeability	45 to 65 acfm/ft^2 @ 1/2 in.H_2O
Weight	10.5 Oz/yd^2
Air to Cloth Ratio, 6 Compartments	2.8 $\frac{acfm}{ft^2}$
Air to Cloth Ratio, 5 Compartments (cleaning or maintenance on one compartment)	3.35 $\frac{acfm}{ft^2}$
Pressure Drop (normal)	4.5 in.H_2O
Pressure Drop (cleaning)	up to 6 in.H_2O

application. "O" rings are used under direct compression for a positive gas seal of the plates containing the jets.

The particulate matter was collected on lightweight stainless steel collection discs coated with Apiezon. The detailed procedures are discussed by Ensor et al. [1976]. The material lost to the walls of the impactor was brushed onto the appropriate substrate surface. A Cahn 4100 balance was used to weigh the discs to 0.01 mg.

Submicron Particles. The submicron particles were measured with an extractive sampling system consisting of a

- Impactor in-stack precutter
- Three-stage dilution system outside of the stack
- Diffusion battery/condensation nuclei counter (Gardner Associates, Inc.)
- Model 3030 Electrical Aerosol Size Analyzer (EASA) (Thermo-Systems, Inc.).

The system is described in detail by Ensor et al. [1976].

Submicron particle testing was a separate and complex task at the inlet and outlet at ports adjacent to the ports used for the cascade impactor tests. The diffusion battery was used at a given location for an entire day. The EASA was moved between the inlet and outlet sampling port as many times daily as feasible.

Dilution and drying of the inlet sample were necessary; dilution ratios ranged from 20X to 100X. Low particle concentrations at the baghouse outlet prevented sample dilution, which meant the dilution system was required only at the inlet sampling site. However, without dilution at the outlet, condensation in the sampling train became a major concern. To reduce the moisture in the sample, two in-line diffusional dryers were used in the sampling train.

An impactor precutter was used at both the inlet and outlet sampling locations. The impactor offered the advantage of a well-defined cut-off at 2.7 microns.

B. Gas Sampling

Gas samples were taken at the inlet and outlet of the baghouse. The water was condensed at the stack and the dry gas piped to the trailer with a diaphragm pump.

A Theta Sensors, Inc., three-gas analyzer (O_2, SO_2, and NO) and a Horiba CO analyzer were used in the test. The inlet and outlet points were alternately sampled every 1-1/2 hours. The instruments were zeroed and calibrated with span gas twice daily.

C. Opacity

The Plant Process Visiometer (PPV) is a light scattering instrument developed at MRI for real-time monitoring of stack opacity. Flue gas is isokinetically withdrawn from the stack with a stainless steel probe and transported into the measurement chamber. The aerosol particles in the chamber were illuminated by a flash lamp with an opal glass filter. The scattering light was detected by a photomultiplier tube at approximately right angles to the flash lamp. The optics have been designed so that the output of the photomultiplier tube is proportional to the extinction coefficient due to scattered light. The instrument is a physical analog of the following equation:

$$b_{scat} = 2\pi \int_0^{\pi} \beta(\theta) \sin\theta d\theta \qquad (1)$$

where

b_{scat} = the scattering coefficient due to scattered light
$\beta(\theta)$ = volume scattering function
θ = scattering angle

If there is no light absorption, the scattering coefficient is identical to the extinction coefficient. The extinction coefficient is related to plume opacity with the Bouguer Law.

$$\text{Opacity (percent)} = \left[1 - \exp(-b_{ext} L)\right] 100 \qquad (2)$$

where
b_{ext} = extinction coefficient, m^{-1}
L = stack diameter, m

The internal calibrator is an opal glass lens of known scattering coefficient. The lens was mechanically placed in the view of the detector for calibration and was retracted into a sealed chamber between calibrations. The PPV was described in detail by Ensor et al. [1974].

The use of the PPV allowed the measurement of opacity at both inlet and outlet of the baghouse. The large difference between inlet and outlet particulate concentration required ranges of 0 to 0.2 m^{-1} and 0 to 5 x 10^{-3} m^{-1}, respectively. The sensitivity of the PPV allowed the measurement of opacity variation of the invisible outlet plume.

D. Test Locations

The inlet sampling location for manual testing was a 6 ft. x 4.5 ft. rectangular duct. Six 3-inch ports offered access to the duct. This location was poor because it was only about 1.4 diameters downstream of a bend, and 1.8 diameters upstream of the baghouse entrance.

The outlet sampling location was at the 5.5 ft. diameter stack. The ports were about 8.4 diameters downstream of the ID fan and about 8 diameters upstream of the stack discharge.

III. Results

A. Overall Collection Efficiency

The primary determination of overall collection efficiency was from the catch in the cascade impactors. Four inlet impactor tests in different ports and two outlet tests were performed each day. The Phase I tests were conducted from September 12 to 26, 1975, and the Phase II tests were conducted from November 2 to 15, 1975. No problems with broken bags were experienced during the tests. Only the particulate results will be summarized here. The details of the test results are reported by Ensor et al. [1976]. The primary manipulated parameter was the generator load of the power plant. The effect of load (air to cloth ratio) is shown in Figure 2. The particle penetration increased rapidly as the air to cloth ratio increased. With increasing load, the baghouse cleaning cycle increased in frequency, and the pressure drop also increased. As indicated at 6 Mw, the baghouse operated during a full day test period without requiring cleaning, and the pressure drop was nearly constant at 3 in H_2O. As the load increased, the cleaning became more frequent. (The average air to cloth ratio is somewhat larger than indicated because of the frequent cleaning.) The large variation in the 12 Mw results is believed to be related to bag cleaning and to the condition of the bag filter cake.

In Table 2, the overall results of the study are summarized. Of particular interest is estimation of the emission rate for the facility. The unit is well within the New Mexico particulate standard of 0.05 lb/10^6 Btu total and 0.02 lb/10^6 Btu for particles less than 2 μm.

The stack was clear during both test periods. The average opacity measured with the PPV was well below the visual threshold of about 2%.

B. Fractional Penetration Results

The major objective of the field test program was to obtain the particle size dependent penetration through the baghouse at various boiler loads. The results are presented in detail by Ensor et al. [1976].

The penetration for an 11 Mw load is shown in Figure 3. The penetration for particle sizes from 0.01 to 1.0 μm were obtained using an Electrical Aerosol Size Analyzer (EASA), and the penetration for particle sizes from 0.5 to 10 μm were obtained with cascade impactors. The testing schedules and process variation are illustrated in Figure 4 with strip charts of the Plant Process Visiometers and CO analyzer and bar charts of the manual tests. The submicron tests were conducted between bag cleaning cycles while the outlet impactor tests included at least three cleaning cycles. The difference between the EASA penetration curve and impactor penetration curve may be due to the increased emissions during the cleaning cycle.

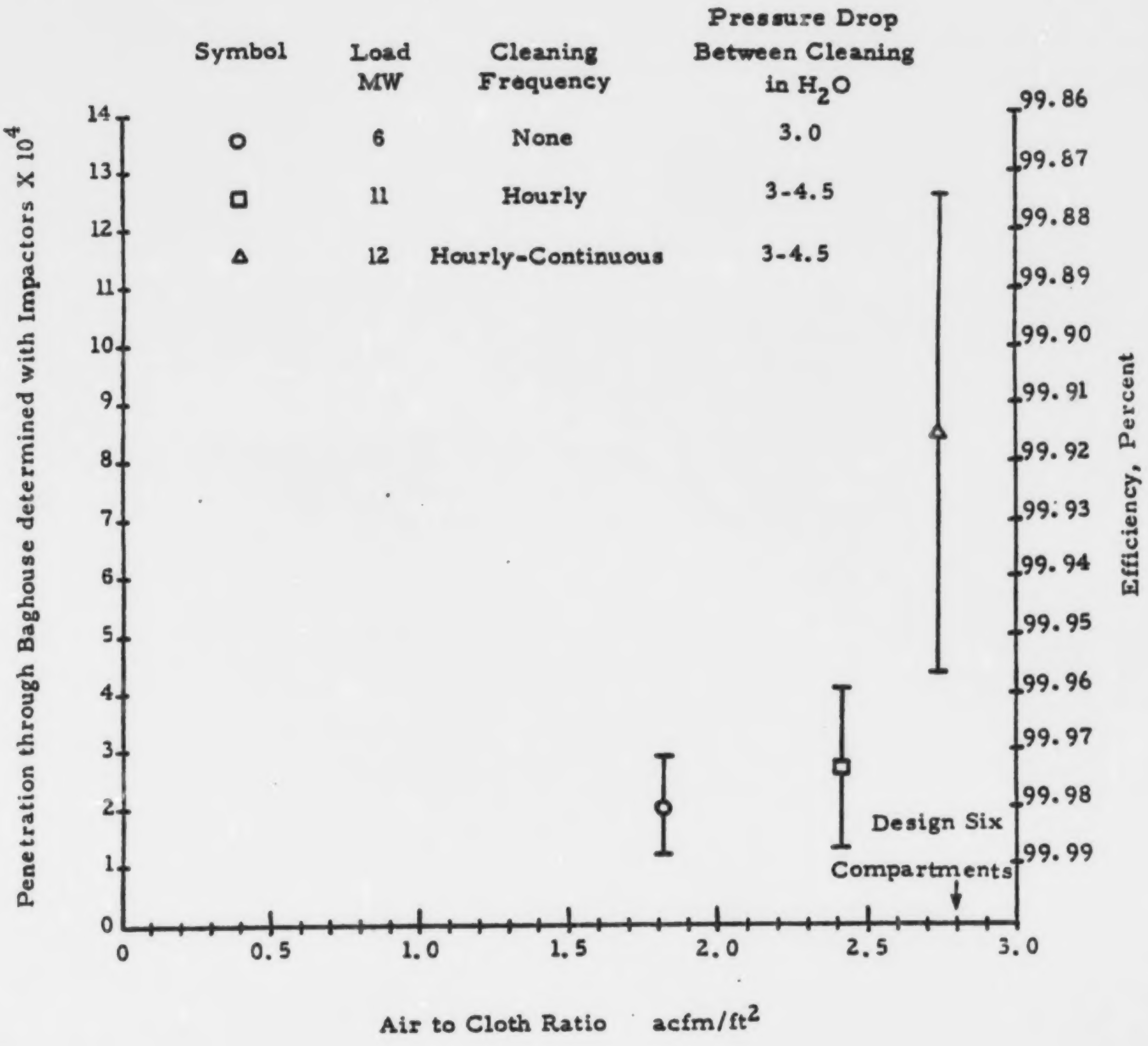

Figure 2. Penetration as a function of Air to Cloth Ratio with one Standard Deviation limits

Table 2. Summary of Overall Efficiency as Determined by Cascade Impactors

Load Mw	Air to Cloth Ratio $ft^3/min/ft^2$	Fractional Penetration (1 Std Deviation)	Number of Days	Efficiency Percent	Outlet Concentration gr/ft^3 (b)	Outlet Opacity Percent	Total(a) Emission $lb/10^6$ Btu	Emission(a) Less than 2 μm $lb/10^6$ Btu
6	1.87 ± 0.09	$(2.07 \pm .85) \times 10^{-4}$	3	99.98	0.00021	~0.02	0.007	0.003
11	2.47 ± 0.07	$(2.74 \pm 1.38) \times 10^{-4}$	9	99.97	0.00087	~0.03	0.02	0.004
12	2.74 ± 0.17	$(8.52 \pm 4.13) \times 10^{-4}$	7	99.92	0.0011	0.6	0.04	0.006

(a) -- Preliminary Results

(b) -- Dry Standard, 760 mm Hg 21.1°C

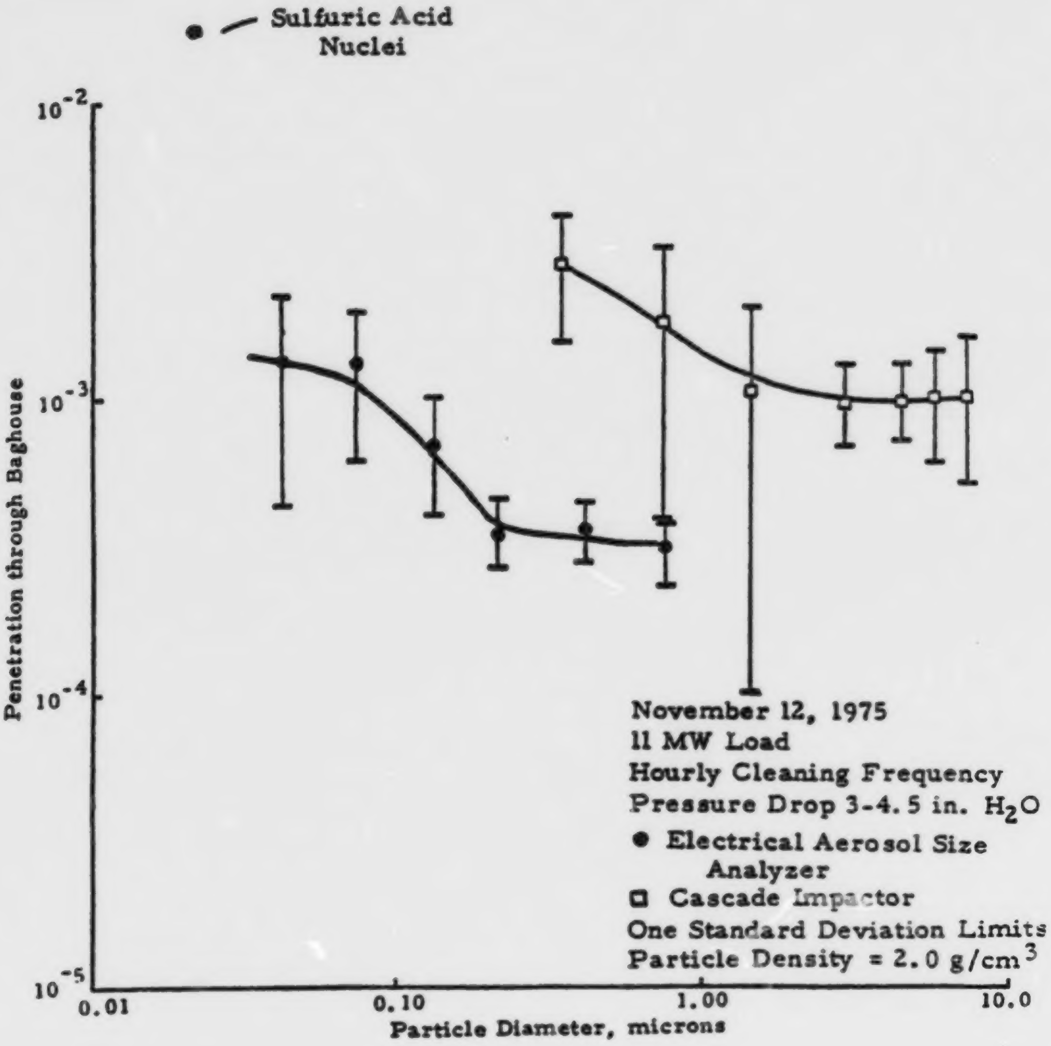

Figure 3. Fractional Penetration through Baghouse,

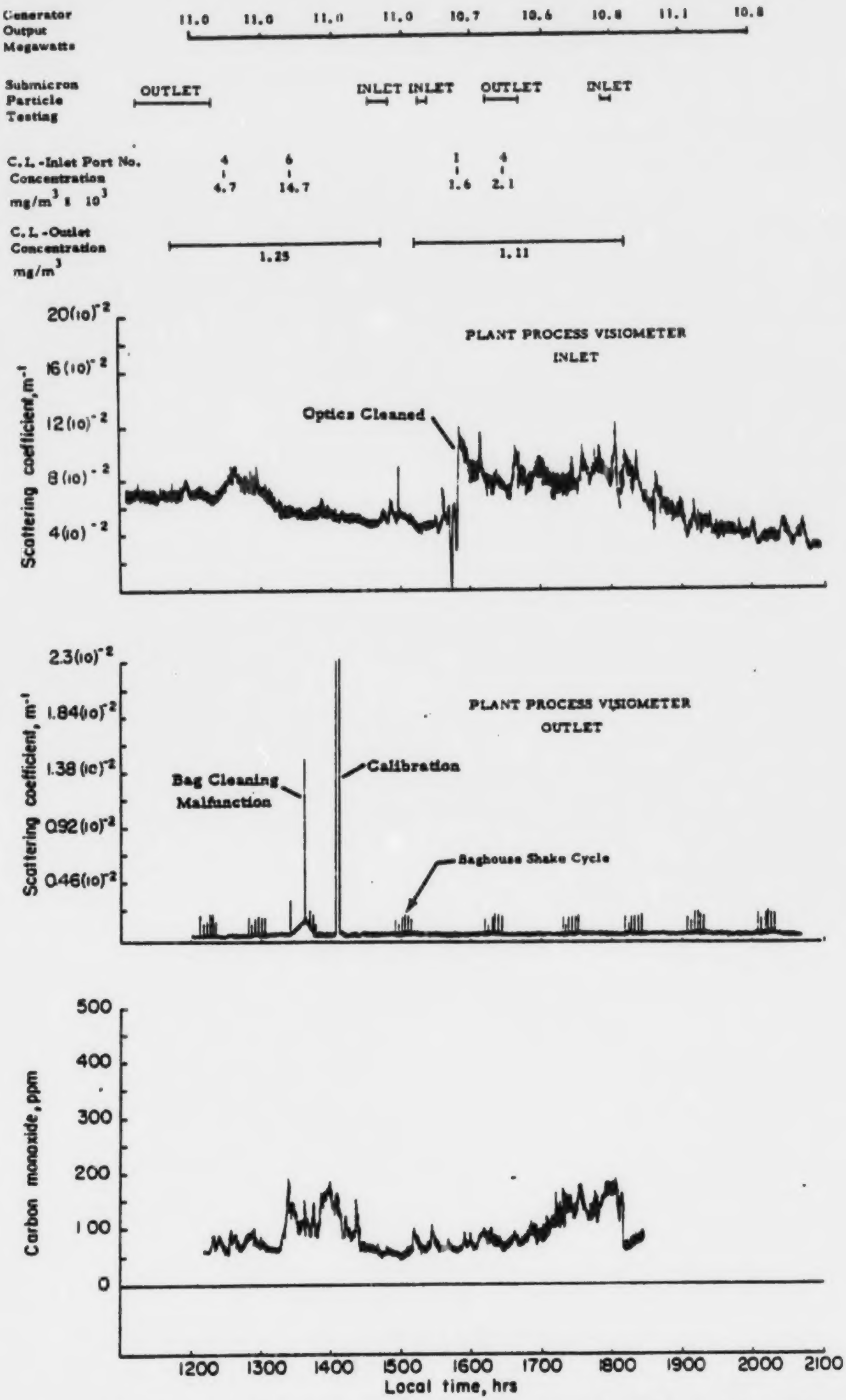

Figure 4. Process Variation During November 12, 1975

The flat penetration of particles greater than 1.5 microns in diameter is illustrative of particle "seepage" through the bags. This "seepage" occurs after the baghouse is cleaned and is a result of particles sifting through the newly cleaned bag until a cake is again formed to aid the filtration.

The increased penetration for the 0.01 μm particle is believed to be due to the formation of sulfuric acid nuclei. The stack temperature was at 210°F which is below the acid dew point. Analysis of cascade impactor final filters indicated a large enrichment of sulfur. The location of the acid mist formation is suspected to be in the uninsulated steel stack or in the outlet duct work in the baghouse. The sulfuric acid nuclei are of very little significance from a mass emission standpoint.

The penetration results, obtained with cascade impactors by Bradway and Cass [1975] the previous year for the same baghouse, were about 10 times greater than obtained in the current study. It is suspected that their data was strongly influenced by bag leakage.

The particle diameter fractional penetration for the half load of 6 Mw is shown in Figure 5. During the test day, the bags were not cleaned. The good match of the EASA and cascade impactor penetrations was due to the lack of bag cleaning cycles.

IV. Engineering Analysis

The objectives of the engineering analysis were to assemble information for both capital and operating costs, determine reliability and identify any major problems. This information is specific for the Nucla installation only, and careful analysis would be required for extrapolation to other sites with different boilers and coal. The analysis was performed considering all three baghouses as a single unit. For detailed cost breakdown, the final report for the Nucla study by Ensor et al. [1976] should be consulted.

A. Capital Costs

The capital costs were estimated from the records of Colorado Ute, Jelco, Inc. (constructor), and Stearns-Roger (engineer). The Nucla plant was retrofitted with baghouses and other additional equipment, thus some of the costs may be unique to this site. A summary of the estimate is shown in Table 3. When evaluating Table 3, the following aspects of the installation should be kept in mind:

1. The cost is for a total of three units. A single larger baghouse of equivalent capacity might be installed for less.
2. The generating units are relatively small, 12.65 megawatts each. For larger units the installed cost per kilowatt of generator capacity would be reduced.
3. Booster fans were not required at Nucla. This resulted in a considerable cost saving.

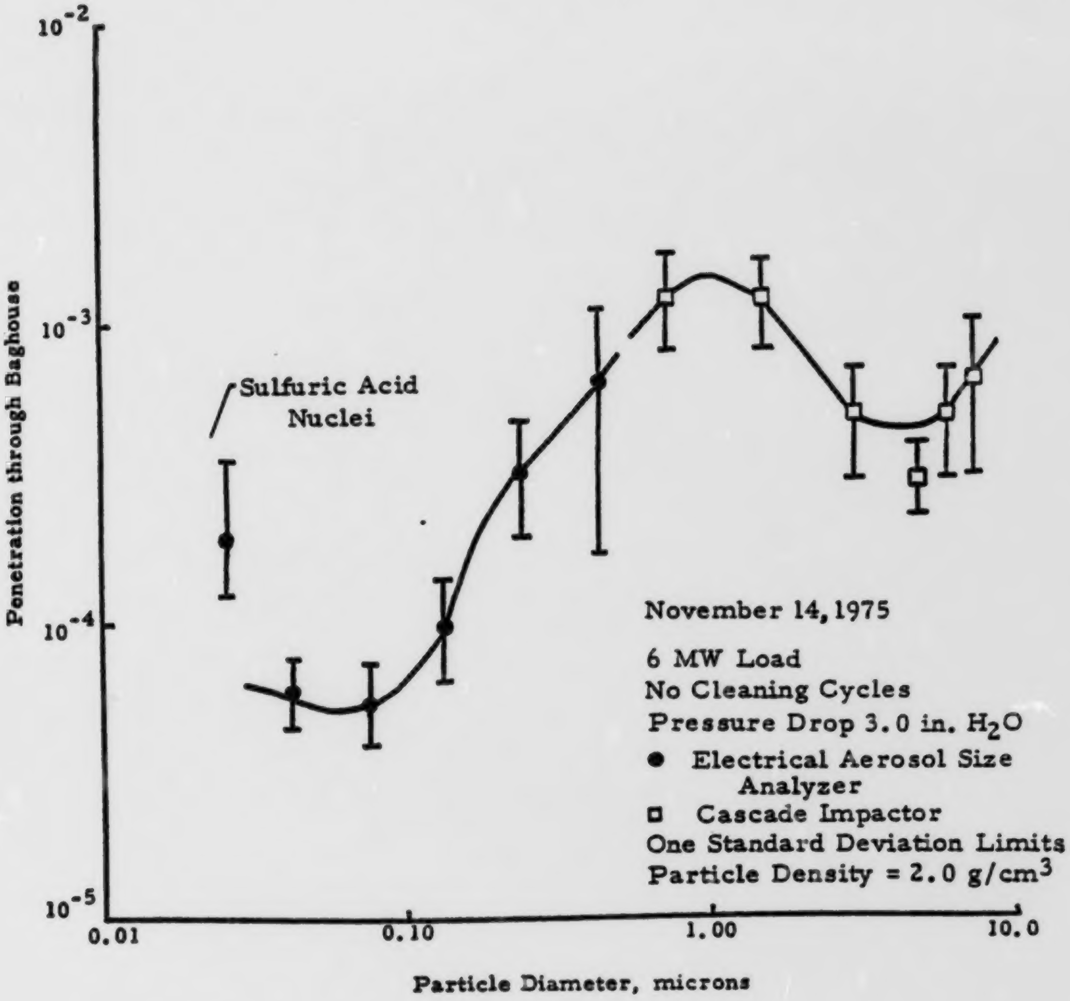

Figure 5. Fractional Penetration through Baghouse,

Table 3. Summary of Capital Cost

		%
Equipment and Installation		
Baghouse and General	$1,740,000	67
Ash Conveyor System	250,000	9
Retrofit Items	210,000	8
TOTAL FIELD COST[a]	$2,200,000	84
Indirect Owner Costs	120,000	5
Engineering and Fee	300,000	11
1973/1974 INSTALLED SYSTEM COST	$2,620,000	100
Estimated Escalation to 1976	680,000	25
1976 INSTALLED SYSTEM COST	$3,300,000	

Unit Factors (1976)
- $87/kw
- $13/acfm
- $36/sq. ft. filter (gross)

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[a] Includes material, labor, supervision, field overhead and constructor's fees.

4. The cost estimate indicated for the "Retrofit Items" is probably low. It does not include the many intangible costs incurred as a result of the retrofit. These costs are reported in the other two categories and cannot be separated because they are very difficult to define. Thus, the fraction of cost attributable to retrofit is probably low.

5. Nucla is a small, remote mountain community. To a large extent, craft labor had to be brought in from Montrose. Nucla has little industry and is not served by rail. These factors undoubtedly increased field costs.

The cost for 1976 was estimated using the projected figures of the Marshall and Swift Equipment Cost Index and the Chemical Engineering Plant Cost Index. The escalation is calculated assuming that the baghouses were installed in the first half of 1976, two years after the actual date.

B. Operating Costs

The operating cost was defined as any additional costs incurred by the utility attributable to the operation of the baghouse. The costs were estimated from Colorado Ute records as well as estimates by plant personnel. The operating costs expected for 1976 are summarized in Table 4.

C. Maintenance

The plant maintenance records were reviewed and summarized in Table 5. The table shows the effort spent for each maintenance category and its trend over the past two years of operation. The figures given in the table are maintenance manhours over the number of occurrences. The four times periods are roughly six months long but have been adjusted somewhat to distribute the 32,600 hours of baghouse operation evenly between the periods. The table shows that during the first three quarters of operation, bag replacement has been the highest maintenance item. In the last quarter, however, bag replacement was minimal and problems with the control system, the dampers and the damper actuators occurred more often.

The trend is also shown in Figure 6, a graph of maintenance manhours required per 1000 hours of baghouse operation. The two lines presented are simply the averages for the previous three or twelve month period. Averaging in this manner tends to smooth, somewhat, the erratic nature of the maintenance requirements. The three month curve is much more responsive and shows a significant drop in March, April and May of 1975. The reduction corresponds with modifications to the cell plate to extend the bag life. During the initial months of operation, it was discovered that severe bag erosion at the inlet of the bags resulted in premature bag failure. During a six-month period starting in September 1974, gas straighteners called "thimbles" were installed at the inlet of the bags. The "thimbles" resulted in a major decline in maintenance.

Table 4. Nucla Fabric Filter System Operating Cost Estimate

Direct Costs	$/year	%	mills/kwh [(b)]
Operation Labor[(a)]	$(9,500)	(3.3)	(0.05)
Maintenance Labor	2,500	0.9	0.01
Maintenance Material	8,500	3.0	0.05
Utilities	31,000	10.8	0.16
Ash Handling	11,000	3.8	0.06
Subtotal, Directs	$53,000	18.5	0.28
Indirect Costs			
Depreciation	127,000	44.3	0.68
Interest	81,000	28.2	0.43
Insurance	3,000	1.0	0.02
Taxes	23,000	8.0	0.12
Subtotal, Indirects	$234,000	81.5	1.25
TOTAL	$287,000		1.53

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[a] Not added since no new costs were incurred.

[b] Based on 188 million kwh/year or 55 percent capacity.

Table 5. Baghouse Maintenance Summary

Maintenance Category	Period[b]				
	Dec '73- July 74	Aug '74- Jan 75	Feb '75- July 75	Aug '75- Dec 75	Total
Bag Replacement	106/24	99/19	46/7	13/4	264/54
Control System	67/15	66/11	22/4	42/10	197/40
Dampers & Actuators	19/6	35/9	20/6	26/7	100/28
Reverse Air Fans	40/7	80/10	10/2	2/1	132/20
Pressure Taps	2/1	23/6	4/1	2/1	31/9
Hopper Heaters	1/1	16/3	14/4	0/0	31/8
Miscellaneous	12/4	7/2	0/0	9/1	28/7
Subtotal	247/58	326/60	116/24	94/24	783/166
Routine	6/2	6/2	6/2	6/2	24/8
Total	253/60	332/62	122/26	100/26	807/174

[a] Units: manhours/occurrences

[b] The four periods have the same amount of baghouse operating time.

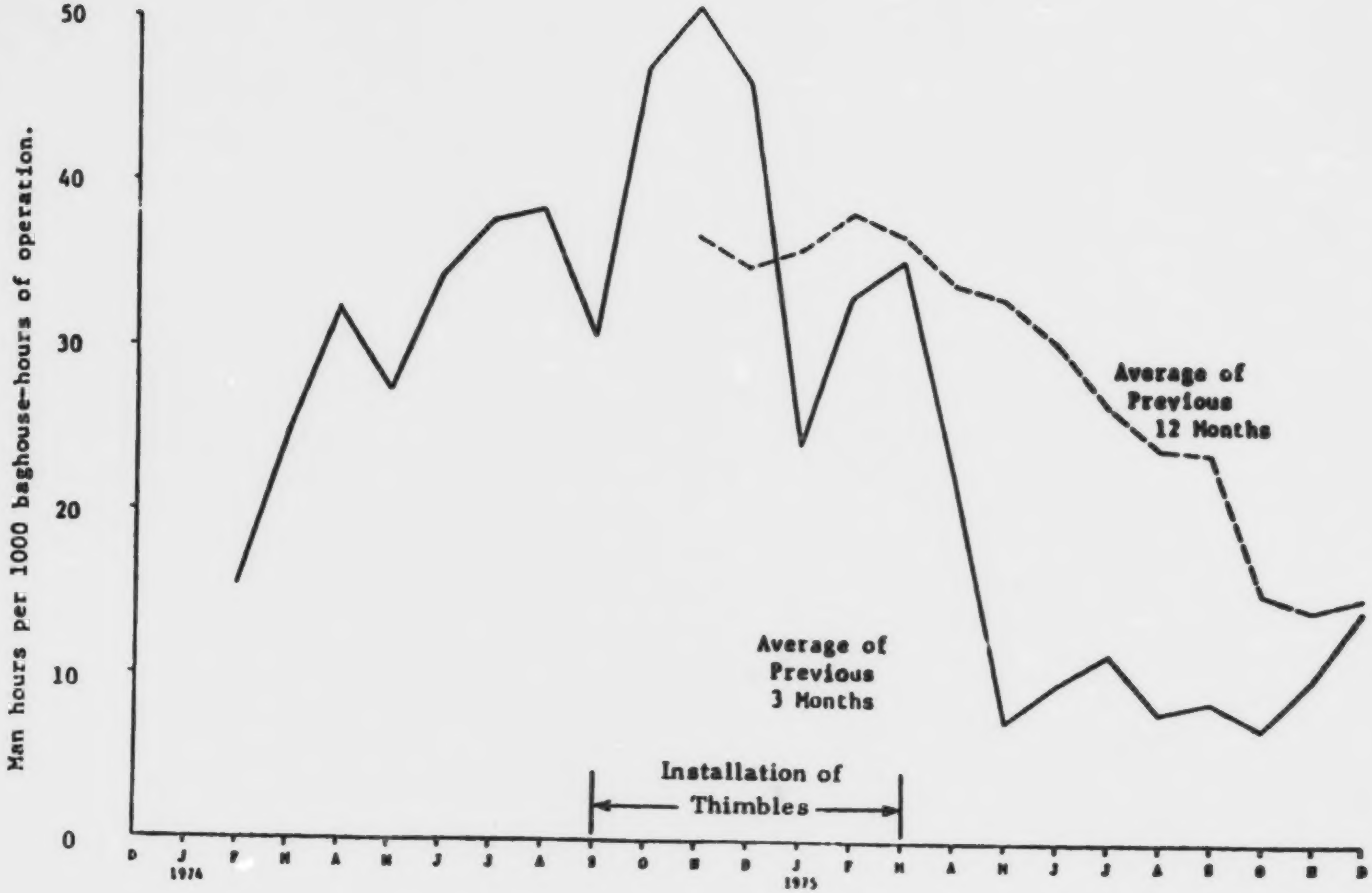

Figure 6. Maintenance Labor Requirements

The curve shows a general leveling of maintenance at about 8-man hours per 1000 baghouse operating hours starting in June. The twelve month curve also shows the definite trend toward reduced maintenance requirements.

This is a presentation of maintenance requirements over the first two years of operation. The system is intended to last twenty years, and equipment that has performed well thus far could cause problems in the future. Projecting maintenance requirements should, therefore, be done with care.

The maintenance requirements for bag replacement have thus far averaged 12 maintenance man-hours to replace 16 bags for each million bag-hours of operation during the first two years of operation. A total of 18 percent of the 2,016 bags were replaced (32,577 baghouse hours of operation). Most of the bags were replaced before thimble installation.

Equipment requiring regular maintenance are:

- the control system
- dampers and actuators
- reverse air for drives
- plugged pressure taps
- hopper heating system
- freezing of compressed air piping

E. Reliability

In addition to good collector design and system design, the reliability of an abatement system is a function of the operating demands and maintenance philosophy of the utility. The units at Nucla are not used for peaking but rather are operated at near full load. This constancy of operation, combined with the attitude of maintenance personnel toward correcting failures as quickly as possible, is to a large extent responsible for the high reliability attained thus far.

Reliability can be expressed in many ways because its definition changes with one's point of view. From the plant superintendent's viewpoint, the reliability is lowered when a situation occurs in which the baghouse either obstructs, impedes or in any way influences the on-stream time of the boiler/turbine. As of December 1975, no such situation has been encountered at Nucla; the station's on-stream factor has been limited not by the baghouse, but by equipment upstream of the baghouse. Thus, the reliability from this point of view is 100%.

A second definition addresses the question of how dependable the units are in producing a clear stack effluent:

$$\text{Reliability} = \frac{\text{(total baghouse operating hours) - (non-clear hours)}}{\text{(total baghouse operating hours)}}$$

Thus far, a bag change has been required in an operating baghouse 50 times. This is used as an estimate of the number of occurrences of a non-clear stack. The duration of each occurrence is the time between when a bag breaks and when its compartment is isolated. When a plume from the stack is noticed, the compartment is isolated within a few minutes. However, considering that bags break during darkness when the stack appearance cannot be observed, the average duration for both day and night is about four hours. Thus, from the standpoint of gas cleaning or stack appearance:

$$\text{Reliability} = \frac{32,577 - (50 \times 4)}{32,577} = 99.4\%$$

It should be mentioned that the high value of this figure is a direct result of the attitude of the Nucla plant personnel toward correcting bag failures as quickly as possible.

Another definition considers the reliability of the six individual compartments in each baghouse. Since startup of the system, all 18 of the compartments have been in service except for 387 hours. This includes approximately 134 hours required for installing the thimbles on the cell plates. From a compartment standpoint:

$$\text{Reliability} = \frac{(32,577 \times 6) - 387}{(32,577 \times 6)} = 99.8\%$$

This may be interpreted to mean that all compartments were in service 99.8% of the time. The major cause of compartment outage has been bag replacement.

V. Summary and Conclusions

The baghouses at the Colorado Ute Electric Association were evaluated with both a field study and engineering analysis. The emissions were within the New Mexico standard of 0.05 lb/10^6 Btu total and 0.02 lb/10^6 of particulate matter less than 2 microns. The stack emissions were not visible during the test. Measurement of opacity with an MRI PPV indicated opacities of less than 1 percent.

The penetration of particulate matter through the baghouse was a sensitive function of the air to cloth ratio. The cleaning part of the operational cycle contributes most of the emissions. The baghouse was an efficient collector of submicron particles with less than a factor of ten variation in penetration as a function of particle size.

The engineering analysis indicated the baghouse had not interferred or limited the operation of the boiler since installation. The maintenance and bag replacement in the unit has been greatly reduced with the installation of "thimble" at the bag entrance.

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OPERATING EXPERIENCE AND PERFORMANCE AT THE SUNDBURY BAGHOUSE

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ABSTRACT

OPERATING EXPERIENCE AND PERFORMANCE AT THE SUNBURY BAGHOUSE

Penna. Power & Light Co.'s Sunbury Plant is located on the Susquehanna river in Snyder Co., Shamokin Dam, Pa. Four of the boilers at the plant are equipped with baghouses that were placed in service in the spring of 1973. The 4 boilers are identical. Each is rated at 415,000 lbs/hour steam flow and has a total gas flow of 220,000 acfm. The boilers use a combination of fuels comprised of a mix of 75% anthracite coal and 25% petroleum coke. The baghouses on each boiler are preceded by mechanical collectors, which have a collection efficiency of 70%.

Each of the baghouses is divided into fourteen compartments with 90, thirty feet long by one foot in diameter bags per compartment. The air-to-cloth ratio is 2 to 1. The bags are cleaned by the reverse air flow method. The initial installation cost, including a new fly ash removal system, was $ 5,500,000. Maintenance and operating cost, excluding complete bag replacement, has averaged $ 30,000. a year.

A total of only 37 bags have been replaced due to failure. Bags in 3 of the 4 baghouses were replaced as a precautionary measure after approximately 2 years of service, at a cost of $ 60,000. per baghouse. The bags in the 4th baghouse are still in service after 3 years of operation. The draft loss across the bags has averaged about 3.0" VWC. The draft loss across the 3 year old bags is presently 3.5" VWC.

A test to determine the removal efficiencies at various particle sizes was performed on 1 of the baghouses in early 1975. Mass efficiencies and efficiencies by particle size were found using mass trains, impactors and diffusional sizing equipment. Some difficulties were met with impactor substrates and with the diffusional apparatus and are described in the paper. Mass efficiencies were 99.88% and 99.93% respectively for new bags and for bags with about 2 years of prior service. Average mass efficiency in the particle size range 1 to 10 m was not lower than 99.5% and reach a peak of about 99.9% at 4μ m. Limited diffusion equipment results indicated an efficiency on the order of 93 to 99% for all particles in the range of about 0.003 to 0.4μ m (by number), and that no particles smaller than about 0.015μ m were present in the baghouse outlet.

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5-18-76

OPERATING EXPERIENCE AND PERFORMANCE AT THE SUNBURY BAGHOUSE

PLANT - GENERAL DESCRIPTION

The Sunbury Steam Electric Station of Pennsylvania Power & Light Company is located on the west bank of the Susquehanna River in Shamokin Dam, Pennsylvania. It generates 402 MW of electricity from six pulverized-coal-fired steam generators and four turbine-generator sets. Turbines Nos. 1 and 2 are each rated at 87.5 MW. Steam is supplied to these turbines through a header system from four Foster Wheeler anthracite fired boilers. (A Foster Wheeler bituminous fired boiler and a 110 MW turbine generator set are also connected into this header system.) Each of the four boilers is rated at 415,000 lb/hour steam flow at 955°F 1350 psig and was placed into commercial service in 1949. The boilers burn a mixture of anthracite (silt and No. 5 buckwheat), petroleum coke, and bituminous coal. (Typical analysis of the plant coal is given in Table 1.) The mixture varies considerably according to the availability and the quality of the fuel. The normal mix is 75% anthracite (consisting of 90% silt and 10% No. 5 buckwheat), 20% coke and 5% bituminous. The normal fuel consumption is 25 tons/hour/boiler. No. 2 fuel oil is used for boiler fire-ups and to stabilize ignition.

PARTICULATE REMOVAL EQUIPMENT

Original particulate removal equipment consisted of a mechanical-electrostatic precipitator set. Particulate removal efficiency with this equipment was 90%. In 1973 baghouses were installed on each of the four anthracite fired boilers. The electrostatic precipitators were gutted due to a weight consideration. The weight reduction allowed utilizing some of the existing precipitator steel to support the baghouse. The mechanical collectors were modified by replacing the original vanes of the multi-cone collectors with new, less efficient vanes. The effect of this was a reduction of mechanical collector efficiency from 78% to 70% while the pressure drop decreased from 4.5 in. VWC to 1.5 in. VWC. The mechical collectors were left in to maintain a lower dust loading on the induced draft fans. It is not certain to what degree, if any, they improve baghouse performance.

The baghouses are installed in series with the mechanical collectors. The flue gas, after leaving the boiler economizer, passes through regenerative air heaters, mechanical collectors, the induced draft fans, and the baghouse. The then clean flue gases discharge to atmosphere through a stack, 300 feet above the ground. (Figure 1)

The baghouses were constructed by Western Precipitator Company. They were initially put into service on the following dates:

No. 1A Baghouse - February 10, 1973
No. 2B Baghouse - March 5, 1973
No. 2A Baghouse - April 14, 1973
No. 1B Baghouse - April 26, 1973

BAGHOUSE - GENERAL DESCRIPTION

Each baghouse is approximately 80 feet long, 40 feet wide and 62 feet high including the dust hoppers. Each baghouse consists of 14 compartments arranged in two rows of seven compartments each. Each filter bag compartment is 19' - 9 in. wide x 10 feet - 1-3/4 in. long x 40 feet high. The baghouse enclosures, including the interior partitions, are constructed of 14 gauge mild steel and are of all welded construction. The 14 gauge partitions and welded construction were decided upon to insure gas tight construction to permit the safe entering of isolated compartments for routine inspections and minor maintenance while the baghouse is in service.

The entire baghouse including the hoppers, collapsing flues and inlet/outlet flues up to the point of tie in with the existing breeching are completely heat insulated and aluminum lagged. The bag filter dust hoppers, located below the thimble floor, are contained in an enclosure constructed of 0.32 in. thick aluminum siding over steel girts. The floor in the hopper enclosure is checker plate in order to contain any ash which may spill during maintenance periods or should malfunctions occur in the ash removal system.

Access for maintaining the filter bags is accomplished in the following manner. A 2 foot wide x 5 foot high insulated compartment access door is provided at the thimble floor elevation for each compartment. A checker plate walkway constructed of 3/16 in. steel plate is provided through the center of the filter bag rows (three rows of 15 filter bags each are located on each side of the walkway) to facilitate changing filter bags as well as permit inspection to detect broken bags. A 2 foot wide x 5 foot high insulated compartment access door and grating walkway are provided in each compartment for access to the filter bags at the bag tensioning framework elevation.

An inlet flue and a collapse air flue run the length of each baghouse between the rows of hoppers. Side ducts from each flue to the individual compartments are provided to the upper portion of the compartment ash hoppers.

The gas inlet damper and the collapse air damper, both timer-operated, are located in the gas inlet flue and the collapse flue, respectively, to permit cycling the compartments through the in-service and cleaning cycles. There is one main collapse fan for each baghouse and two spare collapse fans, each serving two baghouses.

The outlet flue surmounts the baghouse and ties into the existing stack breeching. Each compartment has a manually operated outlet flue damper which permits isolating a compartment during maintenance periods. (See Figure 2 for the baghouse's general arrangement.)

Vent stacks were installed on the top of each baghouse compartment in 1974 and 1975. Originally isolating a baghouse compartment was difficult. Since the gas inlet damper, a leaf type damper, does not provide a positive shut off, too much heat and gas leaked into the compartment. With a portable 7,000 CFM fan mounted on the vent stack enough ventilation is provided to permit entry into the compartment for maintenance with the remainder of the baghouse in service. The vents are 2 feet x 2 feet x approximately 4 feet high, constructed of 10 gauge mild steel and are covered with 2 in. thick insulation and 0.032 in. thick aluminum siding. A spring clamped cover is mounted on top of each vent stack. (Figure 3)

Filter Bags

The design filter ratio was 1.919:1 CFM/square feet of bag and the active filter ratio (based on one compartment being out of service at all times) was 2.067:1. Each baghouse compartment contains 90 bags for a total of 1,260 bags per baghouse giving a total filter area of 115,668 square feet.

Each filter bag is 11-1/2 in. in diameter by 30 feet-4 in. length. The filter bags, Menardi Southern Company's style No. 601T, are made of glass fiber with Tuflex finish (Teflon base, specifications are listed in Table 2). The Teflon coating acts as a lubricant between the glass fiber filaments. It prevents the glass fiber filaments from abrading one another, and thus lengthens bag life. Each filter bag contains seven equally spaced anti-deflation rings. With the rings preventing the bag from collapsing completely during the cleaning cycle, the dust cake is released freely and drops to the hopper below. A metal band is sewn into the top and bottom of the bags for installing the bags. The bags are suspended from a framework in the top of each compartment housing and approximately 50 lb. tension is applied.

Cleaning Cycle

The bags are cleaned by reversing the gas flow through a compartment This partially collapses the bags; some of the dust cake is then released and falls to the hopper below. A collapse fan (12,500 CFM @ 300°F and 10 in. VWC Buffalo Forge Company size 60, 1300 rpm) provides the force for the reverse gas flow. The collapse air flow is drawn through the collapse duct, into the collapse fan, and then is discharged into the baghouse inlet flue where any entrained fly ash is filtered by the bags. (Figure 2)

Each compartment is cleaned in the following manner:

1. The gas inlet damper to the compartment closes, shutting off the flow of "dirty" flue gas to this compartment.
2. The collapse damper opens, allowing a reverse flow of "clean" flue gas from the outlet flue to be pulled through the bags, which partially collapses, and thus cleans the bags.
3. The collapse damper closes.
4. The gas inlet damper opens, returning the compartment to the filtering mode. (Figure 3 illustrates a compartment in the filtering mode and a compartment in the cleaning mode.)

This sequence is repeated on each of the 14 compartments. Twice during the 14 compartment cycle the collapse duct is swept clean. This is accomplished by opening a sweep valve located at the end of the collapse duct (Figure 2). This allows ambient air to be drawn through the collapse duct and fan; and thus, purges any fly ash accumulation.

The time sequence of the cleaning and sweep cycles is detailed in Table 3. There are null periods between the operation of the gas inlet and collapse air damper. This is done to prevent reentrainment by allowing for some settling time for suspended fly ash; and also, to prevent a sudden reversal of air flow through the bags, which would put undue stress on the bags. The air-operated dampers are provided with a slow opening and closing device which also prevents any additional stress on the bags.

Instrumentation

Two instruments, a Bailey bolometer and a differential pressure recorder, are used to monitor baghouse performance continuously.

The bolometer consists of a light bulb and a receiver located in the breeching to the stack on each boiler. The receiver output varies as the light intensity striking it varies, and thus is an indication of smoke density. A 24-hour recorder located in the plant monitors this signal. A reading of zero of scale indicates no smoke and a reading of 100% of scale indicates 100% smoke. The bolometers are not calibrated and just give a relative indication of smoke density.

The differential pressure across each baghouse is recorded on a 24-hour recorder located in the plant control room. The taps are located in the baghouse inlet and outlet flues. An audible alarm is sounded if the differential pressure rises 1 in. VWC above its normal value, giving the operator time to take corrective action.

The differential pressure across each compartment is not monitored. It was decided that a continuous monitoring system would be too costly and of questionable value. There are taps across each compartment but no instruments were installed to indicate differential pressure.

Controls and Alarms

The four baghouse control panels are located in the plant mill room. Each control panel houses two cleaning cycle timers (one of which is a spare), a stepping switch, cycle timer selector switch, collapse fan selector switch, control power "off-on" switch, and the compartment's status indicating lights.

The cleaning control timers and the stepping switch control the cleaning cycle operating the air-operated baghouse gas inlet and collapse air dampers by activating the appropriate solenoid valves. The cleaning cycle control timers were originally electro-mechanical devices. Solid-state electronic timers were added in 1975 and 1976 to facilitate adjustment in the cycle time and increase the reliability.

The compartment status lights indicate the positions of the baghouse gas inlet and collapse air dampers. These indicate whether a compartment is in the filtering or the cleaning mode. Indication is accomplished through the use of limit switches on the dampers.

Adjacent to the baghouse control panels are alarm panels with alarms to indicate a collapse fan failure or a cleaning cycle timer failure.

Baghouse Vacuum Cleaning Systems

Individual vacuum cleaning systems are provided for each baghouse. Vacuum connections have been provided in each bag filter hopper enclosure as well as outside the compartment access doors at the thimble floor elevation.

Baghouse Ash Removal System

The two fly ash removal systems, one per unit, are United Conveyor Corp. vacuum type which pneumatically convey fly ash from the ash hoppers. Air flow required for conveying the ash is produced by an 8 foot Hydroveyor exhauster. The ash/air mixture is transported by the conveyor to the Hydroveyor exhauster where the mixture is combined with water to form an ash slurry. This slurry is discharged into an air separator tank where the conveying air is vented to atmosphere while the ash slurry discharges by gravity into the Units Nos. 1-2 slurry tank.

High pressure (300 psi) water at a rate of 750 GPM is provided at the Hydroveyor for each system by one of three new raw water booster pumps. Plant ash disposal capabilities were increased by the duplication of the existing fly ash disposal system and also the installation of two new 2,000 GPM raw water pumps. The new fly ash disposal system is composed of one ash slurry tank, an ash slurry pump house, six slurry pumps (in series) with 1600 GPM capacity and over 2 miles of 10 in. buried steel pipe through which the slurry is pumped to the fly ash settling basin.

OPERATION

The flue gas enters the baghouse at 350°F and leaves at 310°F. The flue gas temperature normally stays within this range except during boiler startups and shutdowns when the temperatures are in a lower range, anywhere from ambient air temperatures to normal operating temperature. Several times since 1973 the regenerative air heater drives have failed. This has caused the baghouse inlet temperature to rise to as high as 510°F. The high temperatures were sustained for about 1 to 2 hours. There were no discernable effects on the bags.

The design pressure drop across the baghouse and baghouse duct work was not to exceed 6 in. VWC, which was dictated by the capacity of the existing induced draft fans. Average draft loss across the baghouses is as follows:

Baghouse No.	Avg Draft Loss (in. VWC)	Bags Installed
1B	3.5	4/26/73
2A	3.5	11/20/74
2B	3.0	2/23/75
1A	2.5	3/20/75

The baghouses are on the discharge side of the induced draft fan and thus, are under a positive pressure.

The gas flow through each baghouse is 222,000 ACFM at a temperature of 325°F. Baghouse inlet grain loading is approximately 2 grains/cu ft.

There is no visible stack discharge from any of the baghouses, except for vapor condensation during the coldest days of the year (approximately 10°F or lower). When a large bag failure occurs (e.g., a 2 to 3 foot tear in a single bag) there will sometimes be a very faint stack discharge (approximately 5% opacity or less) depending on the location of the bag failure.

Boiler Start-Ups and Shutdowns

To prevent blinding of the bags with the oil used during boiler fire-ups, before a new set of bags is placed in service they are precoated with fly ash remaining in the boiler gas passes. The boiler is brought on the line and the baghouse cleaning cycle is not activated until 1 hour after coal is fired. This allows an additional coating to form on the bags.

Each boiler is shut down and fired-up several times a year. Outside of the precoating of new bags, no special precautions are taken. During normal fire-ups, shutdowns and outages the cleaning cycle is left on. The continued cleaning during outages does not remove the initial fly ash coating but will prevent the formation of a thick cake, which can accumulate from moisture carry-over while water lancing boiler slag deposits.

The number of fire-ups for each boiler is given below.

Boiler No.	1973	1974	1975
1A	6	3	2
1B	3	5	2
2A	3	3	5
2B	7	3	3

Operating Problems

The major operational problem has been occasional baghouse high differential pressure. An alarm sounds to warn the operator of the high differential. The usual causes are either collapse fan failure, cleaning cycle timer failure, or gas inlet or collapse air damper failure. Once the problem is corrected the differential pressure returns to normal in a short time, approximately 30 minutes. Normally the problem is corrected without affecting boiler operation. However, if the differential increases to approximately 5 or 6 in. VWC the boiler steam flow must be reduced due to the loading on the induced draft fans.

Baghouse problems infrequently (about 5 or 10 times/year for all baghouses) have required reductions in boiler load of about 20,000 to 40,000 lb/hr steam flow, but have never forced a boiler off the line.

At present there are no alarms for collapse fan failure. Flow switches were installed in the collapse fan duct to indicate fan failure, but the switches proved to be inoperative. It is planned to install other alarms. At present collapse fan failure is detected from increased differential pressure signals.

When a main collapse fan fails the spare collapse fan is put into service by opening blast gate (butterfly type) dampers. The spare fan is normally filled with fly ash due to leakage past the blast gate dampers, and normally the fan must be cleaned out before it is put into service. This can take as much as 2 to 3 hours. Originally the spare fan was isolated by sliding gate dampers. These dampers provided a tight seal; however, they were difficult to open and close. It took four men with a chain hoist approximately 4 hours. There is a study to determine a method to prevent fly ash accumulation in the spare fan and still allow easy damper operation. As a temporary measure the spare fans are pressurized with compressed air to prevent in-leakage.

If a cleaning cycle timer fails an alarm is sounded and the operator switches to the spare timer. Originally the timers failed frequently. This has been alleviated by adding more reliable, electronic timers.

Damper failures can sometimes be detected by observation of the differential pressure chart. As the dampers open and close the differential pressure swings. If a damper fails, the absence of this pressure swing leaves a "gap" on the differential pressure chart (Figure 4). If a high differential pressure alarms, the dampers are routinely checked for proper operation. The operator must go up to the baghouse and visually observe damper operation through the complete cycle (a total of 32 minutes).

When a baghouse hopper becomes plugged and fills with fly ash, a high pressure spike, in sequence with the 32 minute cleaning cycle, is indicated on the differential pressure recorder. No operational problems are experienced. The operator must sledge-hammer and/or air-lance the hopper to free the blockage.

Since the number of bag failures has been small, they have not caused operational problems. Most failures are single-bag; two-bag failures are the most amount occurring at a single time. Isolating a compartment increases the differential pressure by approximately 0.5 in. VWC. Two compartments have been taken out of service together with an increase of approximately 1.0 in. VWC in differential pressure and no reduction in boiler load was necessary. Any number of compartments may be taken out of service without affecting boiler load as long as the differential pressure remains below 6.0 in. VWC.

There have been two major baghouse pressure excursions. On Baghouse No. 2A in July 1975 after a boiler outage the draft loss increased from 3.0 in. to 6.0 in. VWC over a 3 week period. This was apparently due to moisture carry over to the bags from heavy water-lancing of boiler slag deposits during the outage. Several methods were tried to reduce the draft loss: two collapse fans were placed in service on the baghouse, the collapse dampers were opened and closed several times during each 2 minute compartment cleaning time (this was attempted to dislodge any thick dust cake on the bags), all hoppers were inspected and any blockage was cleared, and the thimble floor was vibrated by pounding it with sledge hammers to dislodge any buildup on the thimbles. (This thimble buildup, a fly ash deposit reducing the 12 in. diameter thimble size to 6 in. to 9in. diameter, has been noted on all baghouses but never has had any noticeable effect on baghouse performance.) All of these methods proved unsuccessful. A compartment was isolated and the bags were inspected. The dust cake on the bags was very thick, from 1/2 in. to 3/4 in. It was decided to manually shake the bags. Each bag was grasped between the palm of each hand and vigorously shaken from side to side. The release of the dust cake could be felt as the dust fell to the hopper below. Alternately, all of the compartments were taken out of service and the bags were manually shaken. When all compartments were returned to service the pressure drop had decreased to 3.5 in. VWC and has remained stable since.

In the fall of 1975 the pressure drop on Boiler No. 1B increased from 3.5 to 5.0 in. VWC over a 3 month period. The bags were inspected during the boiler outage. The cake was at most about 1/4 in., so the bags were not manually shaken. All other methods used on No. 2A were tried with no success. With the new electronic timers it is possible to readily change the cleaning cycle time. On January 23, 1976 the time was reduced from the 30 minute cycle to a 15 minute cycle, doubling the number of collapses. In the last part of February the draft loss began to decrease until it reached 3.5 in. VWC at the end of February. The cleaning cycle was returned to the normal 30 minute cycle on March 16, 1976. The draft loss has remained stable at 3.5 in. VWC since. Because of the time frame it is uncertain whether the 15 minute cleaning cycle caused the draft loss reduction. Normally, when a change is made to reduce draft loss the effect is noticed in several hours. Additional background information on the Sunbury installation may be found in references 1, 2 and 3.

MAINTENANCE

Most maintenance hours have been spent on bag replacement, collapse fan repairs and air-operated damper repairs.

Bag Replacement

The procedure for bag replacement is detailed in Figure 5.

During each annual boiler outage all baghouse compartments are inspected. If an accumulation of dust is found on the compartment floor, each bag in the compartment is inspected for possible failure and all failed bags are replaced.

To replace a bag with the boiler in service, the associated compartments must be isolated by closing the gas inlet and outlet dampers. The lower and upper doors of the compartment are opened to allow ambient air to circulate. The cover is removed from the vent stack, and a portable 7,000 CFM fan is set on the vent stack and started to provide forced ventilation. It normally takes 3 to 4 hours to ventilate the compartment sufficiently for men to enter. It is normally still necessary to wear a mask to prevent inhalation of sulfur dioxide. The compartment can then be entered and the failed bag replaced. The entire procedure(isolating and ventilating the compartment, finding the leak, replacing the bag, and returning the compartment to service)takes approximately 6 to 8 hours.

Detecting Bag Failures

Bag failures are detected by daily observations of the bolometer charts. When a bag has failed the bolometer senses the increased particulate emissions;

generally, there is no visible stack discharge. The bolometer chart will indicate a periodic spike, in sequence with the cleaning cycle. There will be a decrease in the bolometer reading when the compartment with the failed bag is removed from service, a spike when the compartment is returned to service, and a settling out to a higher than normal reading directly after (Figure 6). Some spikes are not easily discernable and a careful study of each chart is necessary. Also some non-periodic spikes occur, but these do not indicate an abnormal condition.

When it is determined what compartment has the failed bag, the compartment is taken out of service. The bolometer readings then return to normal. The compartment is entered, and each of the ninety bags is inspected. An accumulation of fly ash on the compartment floor is a tell tale sign that there is an actual bag failure in the compartment. However, bag failures have been found in compartments with no fly ash accumulations. The bags are inspected by holding a flashlight at the bottom of the bag and shining it up the side of the bag. Any tears are illuminated in this manner. Also, a slight tap on the bag will, if there is a failure, cause a stream of fly ash to flow from the bag which is illuminated by the light; providing an additional check for failures.

The failed bag is replaced, the compartment floor is cleaned, and the compartment is returned to service.

Bag Failure Rate

As of May 1, 1976 a total of 37 bags have failed. A list of the failures is detailed in Table 4.

As a precautionary measure bags in three baghouses were replaced after approximately 2 years of service (dates are listed in Table 4). The 4th baghouse, No. 1B, still has its original set of bags. The bags will remain in service until an increased bag failure rate dictates their replacement.

Maintenance Problems

The collapse fans have been a continual problem. The fans were not designed to handle dirty gas. There is baffling in the hoppers to prevent the collapse fan from picking up dust. However, this has not been effective. The fan casings and blades are worn thin from the abrasiveness of the fly ash. This has caused excessive fan vibration and gas leakage from the fan. Frequent maintenance is required on the fans.

The air operated dampers have caused continual problems. They are located in an environment of high temperatures (150°F to 200°F) and occasional dusting. This results in increased maintenance and occasional damper failure.

ECONOMICS

The initial installation cost of all four baghouse installations, including the fly ash removal system, was $5,500,000. Installation cost and yearly operating and maintenance costs are detailed in Table 5. Increased mechanical and electrical maintenance cost in 1975 was the result of increased collapse fan maintenance and the high pressure drop problem encountered on Baghouse No. 2A.

PERFORMANCE

In order to assess the performance of a fabric filter system as applied to a utility boiler, EPA asked for permission to perform efficiency tests after the Sunbury baghouses had been in service for a few months. Permission was granted and one set of tests was made in August 1973.[4] The results of these tests were impressive and further testing was scheduled to be done by GCA Corporation. The results of the GCA testing, performed in January, February and March 1975, are discussed in this paper.[2]

Of greatest interest to EPA was the performance of the baghouse in the fine particulate range, i.e..below about 3 μm. Along with mass efficiency testing there were also measurements made with impactors and diffusion apparatus. The problems associated with the latter two types of devices led to some rather large error bands, but with over 30 tests made there is reasonable confidence in the results.

GCA performed their tests after the baghouses had been in service for approximately 2 years. Since bag changeover was to coincide with a scheduled outage, this situation allowed testing of bags at the end of their planned service life and also (for the new set of bags) at the beginning of their service life.

Mass Results

For 19 tests made at several combinations of fuel mix and boiler load, the average outlet concentration was 0.0017 grains/dscf or in terms of emission rate: 0.0041 lbs/10^6 BTU. These tests were for the bags which had been in service for 2 years. Based on an average inlet loading of 2.6634 grains/dscf the average efficiency was 99.94%. For the new bags the average outlet concentration was 0.0024 grains/dscf; emission rate was 0.00543 lbs/10^6 BTU; average inlet loading was 2.2853 grains/dscf and average efficiency was 99.89%. The slightly higher outlet loading was consistent with slightly lower pressure drop through the new bags, and the new bags would be expected to have performance similar to the old bags after sufficient stabilization time. The results do indicate that baghouse efficiency does not deteriorate with time, and tends to get better through the service life of the bags.

For all 31 tests the range of outlet loadings was from 0.0013 to 0.0033 grains/dscf (0.0028 to 0.0101 lbs/10^6 BTU).

Impactor Results

Apparent collection efficiency by particle size is shown in Figure 7. For the size range covered, 10 to 0.1 μm, it is apparent that the baghouse maintains nearly uniform, high reduction of particles entering the system. All sizes in the impactor range are collected with better than 99 % apparent efficiency. There is some question as to whether or not particles can change their size as they go through the collection system, but the result at the baghouse outlet is a clear stack.

Condensation Nuclei Counter Results

A condensation nuclei counter (CNC) was used to count numbers of particles in the range from about 0.003 to 0.5 μm. Although the data are sketchy they indicate an apparent collection efficiency of between 93.0 and 99.7 % for particles in this size range. It should be emphasized that this is a number efficiency and not a mass efficiency.

General Comments

It is obvious that the Sunbury baghouses (as do the Nucla baghouses)[5] do a fine job at collecting particulate matter. The GCA testing also revealed that collection efficiency was not significantly affected by changes in fuel composition, boiler load, filtration velocity or cleaning regime within what might be normal limits of change. The baghouse seems to operate like a smoothing device; regardless of inlet variations the outlet showed very little change. For the changes that did occur, regression equations were formulated which showed particle penetration to be primarily a function of inlet concentration, percent moisture, filtration velocity and pressure drop.

Summary of Baghouse Performance

As far as filtration performance is concerned, the Sunbury baghouses do an excellent job. Mass efficiency is 99.9 percent, grade efficiency is 99% or better in the impactor range, and the number efficiency for particles in the CNC range appears to be between 93 and 99.7%. Changes in process and collector operating variables have little effect on outlet concentration.

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2. Cass, R.W. and R.M. Bradway. Fractional Efficiency of a Utility Boiler Baghouse - Sunbury S.E. Station - G.C.A. Corporation NTIS No. (to be assigned). EPA-600/2-76-077a, March 1976.

3. Operating Instruction for Therm-o-flex Filters for Pennsylvania Power & Light Co., Sunbury, Pa., Joy Manufacturing Company, Western Precipitation Division, February 1973.

4. Harris, D.B. and J.H. Turner. Particulate and SO_2/SO_3 Measurements Around an Anthracite Steam Generator Baghouse. Particulate and Chemical Processes Branch, Control Systems Laboratory, Office of Research & Development, Environmental Protection Agency, National Environmental Research Center, Research Triangle Park, N.C. November 8, 1973.

5. Bradway, R.M. and R.W. Cass. Fractional Efficiency of a Utility Boiler Baghouse -- Nucla Generating Plant. GCA Corporation. NTIS No. PB 246 641/AS. EPA 600/2-75-013a.

Table 1. TYPICAL PLANT COAL ANALYSIS

	Anthracite	Bituminous	Petroleum Coke
Heating Value (wet basis) BTU/lb)	8,680	12,159	14,054
% Ash	28.0	15.3	1.7
% Moisture	16.4	5.3	6.4
% Volatile Matter	8.4	27.1	13.3
% Sulfur	0.8	2.6	5.3

Table 2. MENARDI FILTER BAG, STYLE NO. 601T

Fabric	Glass Fiber
Average Permeability	38.5
Mullen Burst Strength (Avg. P.S.I.)	595
Weight (oz. per sq. yd.)	10.0
Thread Count	54 X 30
Weave	3 X 1 Twill
Finish	Teflon 10% by weight
Warp Yarn	150's 1/2
Fill Yarn	Bulked 1/4

Table 3. CLEANING/SWEEP CYCLE TIME SEQUENCE

	Elapsed Time (Sec)
I. Clean compartments (1) through (7) one at a time	
Gas Inlet Damper starts to close	0
Gas Inlet Damper closed	2.0
Collapse Air Damper starts to open	16.0
Collapse Air Damper open	17.0
Collapse Air Damper starts to close	68.0
Collapse Air Damper closed	69.5
Gas Inlet Damper starts to open	81.0
Gas Inlet Damper open	83.0
Next Gas Inlet Damper starts to close	122.0
II. Sweep Collapse Air Duct	
Last Compartment Gas Inlet Damper open	0
Sweep Valve open	40
Sweep Valve closed	120
Next Compartment Gas Inlet Damper open	165
III. Clean compartments (8) through (14) one at a time.	
Repeat cycle shown under I	
IV. Sweep Collapse Air Duct	
Repeat cycle shown under II	

Table 4. BAG FAILURES[a]

Baghouse	1973	1974	1975	To May 1976	TOTAL
1A	5	1	2	2	10
1B	0	6	2	2	10
2A	3	0	0	0	3
2B	4	7	2	1	14
				TOTAL	37

Complete Bag Replacements

1A	3-20-75
1B	Bags never replaced
2A	11-20-74 (except for compartment No. 8)
2B	2-23-75

a. Failures include neither those from faulty installation or damage during maintenance (about 23 bags) nor test bags.

Table 5. INSTALLATION, OPERATION AND MAINTENANCE COSTS

Bag Filter Installation Cost

Expenditure Description	Total Cost ($)
Western Precipitator Contract (4 Baghouses)	3,201,700.
Structures & Improvements	372,100.
Land & Land Rights	1,500.
Boiler Plant Equipment	1,190,000.
Ash Removal Equipment	
Precipitator Modifications	
Accessory Electrical Equipment	72,000.
Overhead	662,800.
Total Construction Cost	5,500,100.

Estimated Operating & Maintenance Cost (@ $12/manhour)

	Cost ($)		
Cost Description	1973	1974	1975
Collapse Fan Power Consumption	18,600.	18,600.	18,600.
Air Compressor Power Consumption	3,000.	3,000.	3,000.
Instrument Department Labor	950.	950.	1,800.
Mechanical Maintenance Labor	2,130.	5,840.	15,678.
Electrical Maintenance Labor	7,410.	3,800.	6,000.
Construction Department Labor	3,950.	2,350.	2,000;
Complete Bag Replacement			
Material ($ 48,000/baghouse)		48,000.	96,000.
Labor ($ 11,000/baghouse)		11,000.	22,000.
TOTAL COST	36,040.	93,540.	165,078.

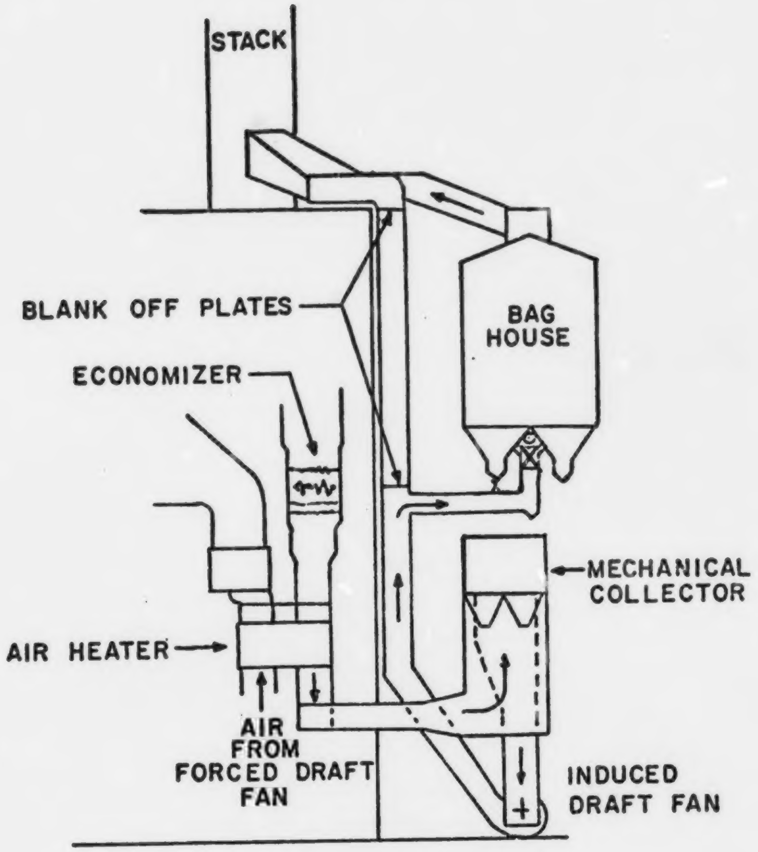


Figure 1. Location of dust removal equipment

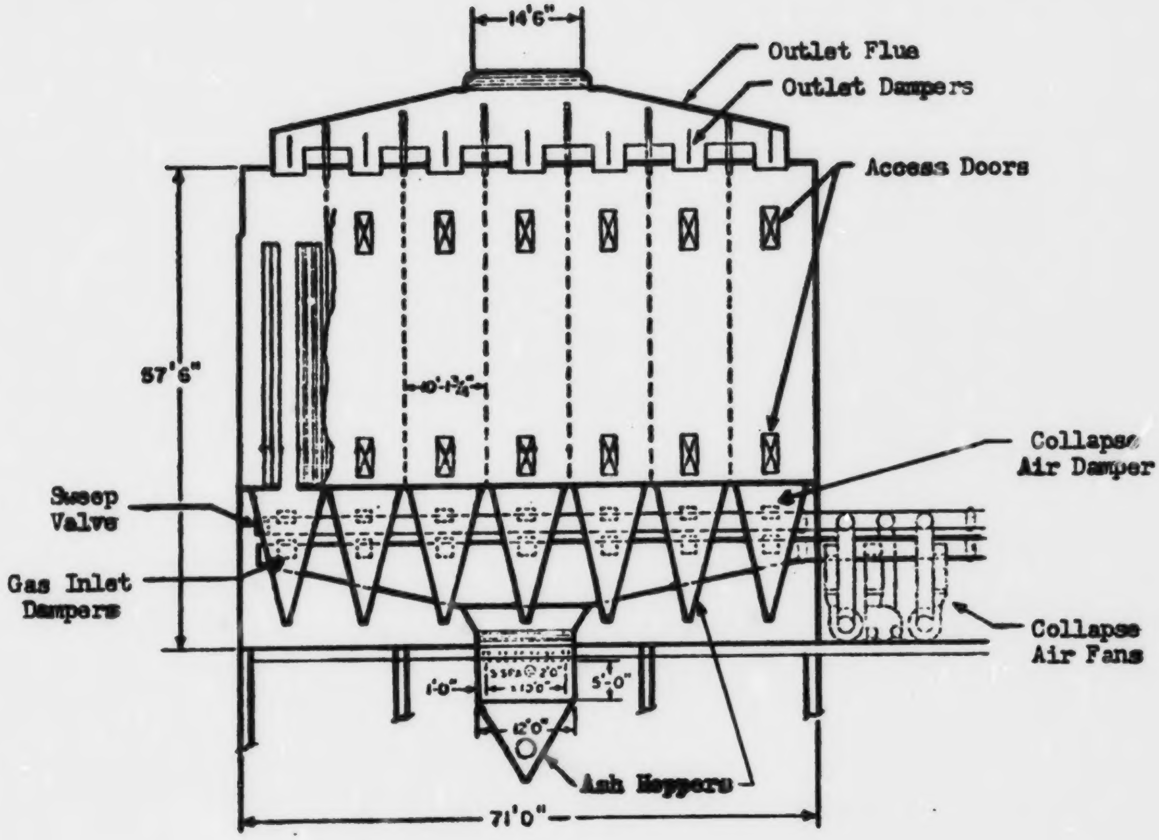
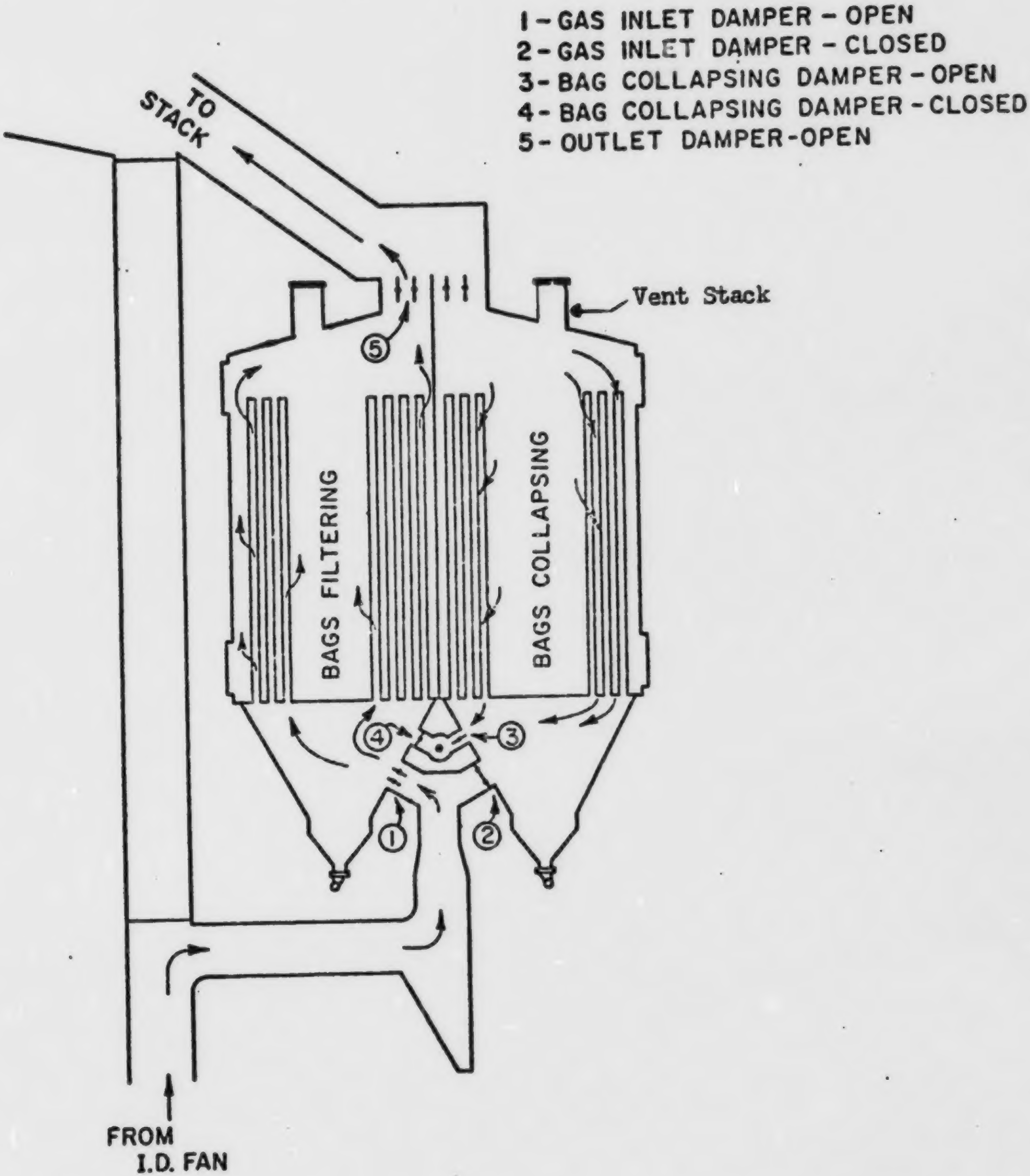


Figure 2 - Baghouse General Arrangement



1-GAS INLET DAMPER - OPEN
2-GAS INLET DAMPER - CLOSED
3-BAG COLLAPSING DAMPER - OPEN
4-BAG COLLAPSING DAMPER - CLOSED
5-OUTLET DAMPER-OPEN
TO STACK
Vent Stack
BAGS FILTERING
BAGS COLLAPSING
⑤
④
③
①
②
FROM I.D. FAN

Figure 3. Gas flow through baghouse compartments during normal operation and cleaning

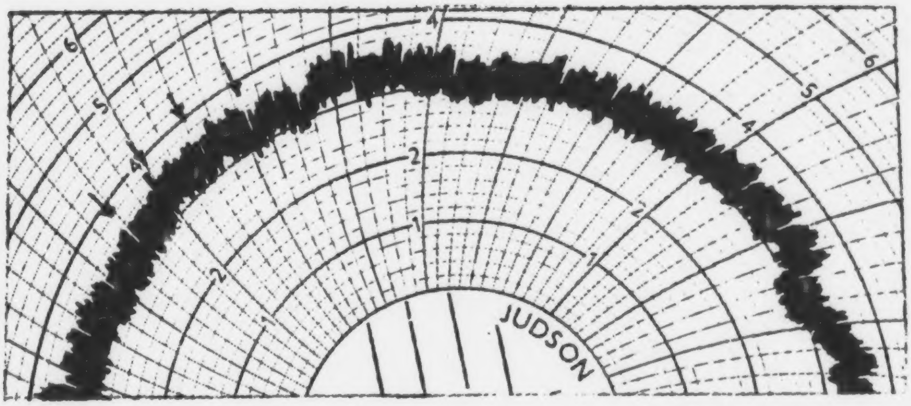
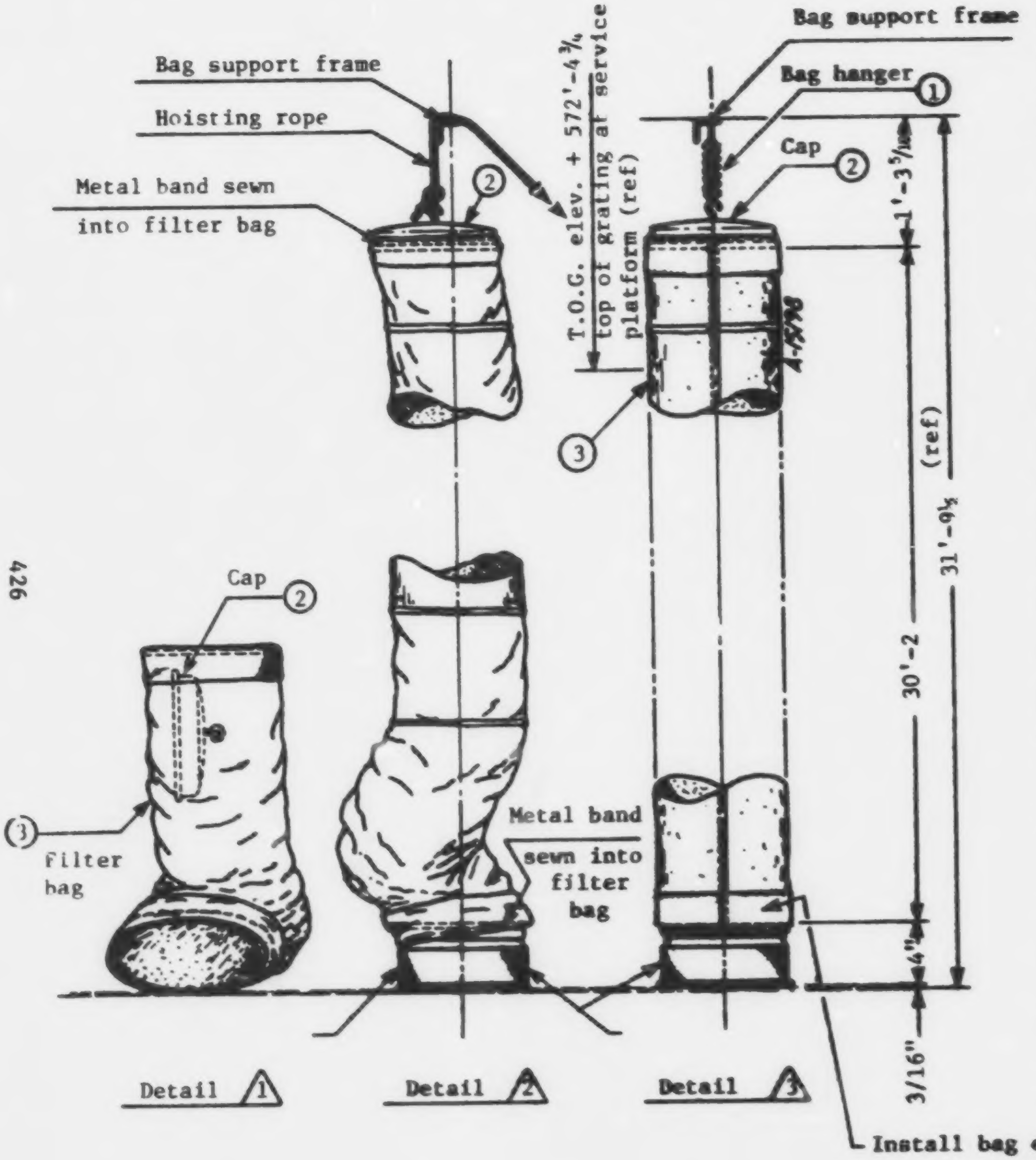


Figure 4. Differential Pressure Chart - indicating a damper malfunction on 2A baghouse (notice arrows).
A normal differential pressure chart has no gaps.



INSTRUCTIONS FOR HANDLING AND INSTALLATION OF GLASS CLOTH FILTER BAGS

Handling

Glass cloth is soft, flexible, and durable. However it is subject to types of damage not always associated with other fabrics. Care in its handling will be repaid in greater efficiency and longer life.

Unpack bags carefully, one by one, as they are installed. Avoid contact of the fabric with projections of metal or rough wood which may cause snags or tears.

Never lay bags on the thimble-plate of the baghouse where they may be walked upon.

The fabric should never be pulled by grasping between thumb and fingers, which may cause slippage of the yarns. Folded bags should not be subjected to heavy pressure.

Installation

Step One. After unpacking, place filter Bag 3 in box on thimble plate near position of assembly. Raise one end of filter bag and insert Cap 2 as shown in Detail 1. Rotate cap 90° and pull upward until metal band sewn into bag cuff fits snugly against rolled rim of cap, as shown in Detail 2.

Step Two. Hook onto Cap 2 with hoisting rope and raise bulk of Bag 3 off thimble plate. Place lower end of bag over thimble, as shown in Detail 2. Push bag end downward until thimble rim is completely covered. Then gently and evenly. Move bag end upward until metal band sewn into bag cuff fits snugly against rolled rim of thimble.

Step Three. Hoist bag to full height and replace rope with Hanger 1 as shown in Detail 3. Tighten hanger to develop 50# tension on bag by compressing hanger spring 1½ inches.

Figure 5. Filter Bag Detail

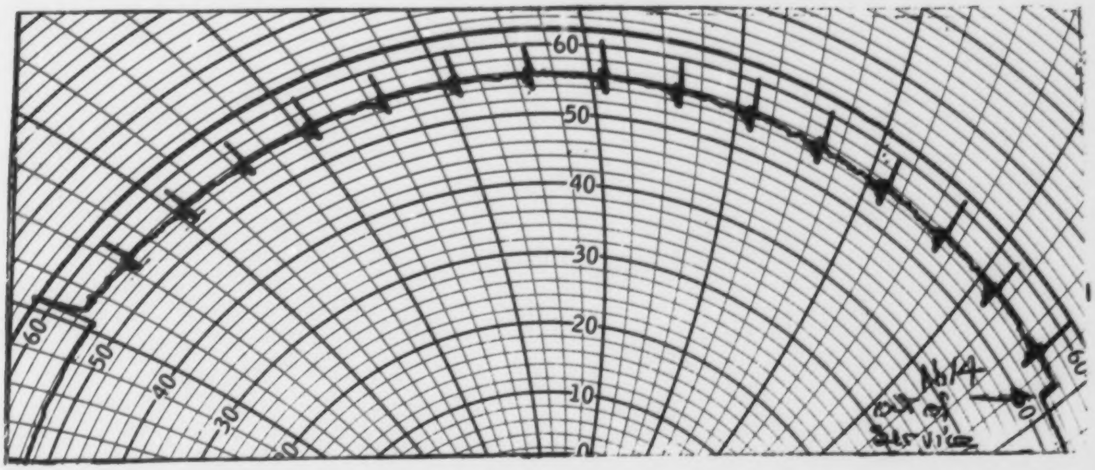


Figure 6. Bolometer Chart - indicating a bag failure in compartment No. 14 of No. 2B baghouse.

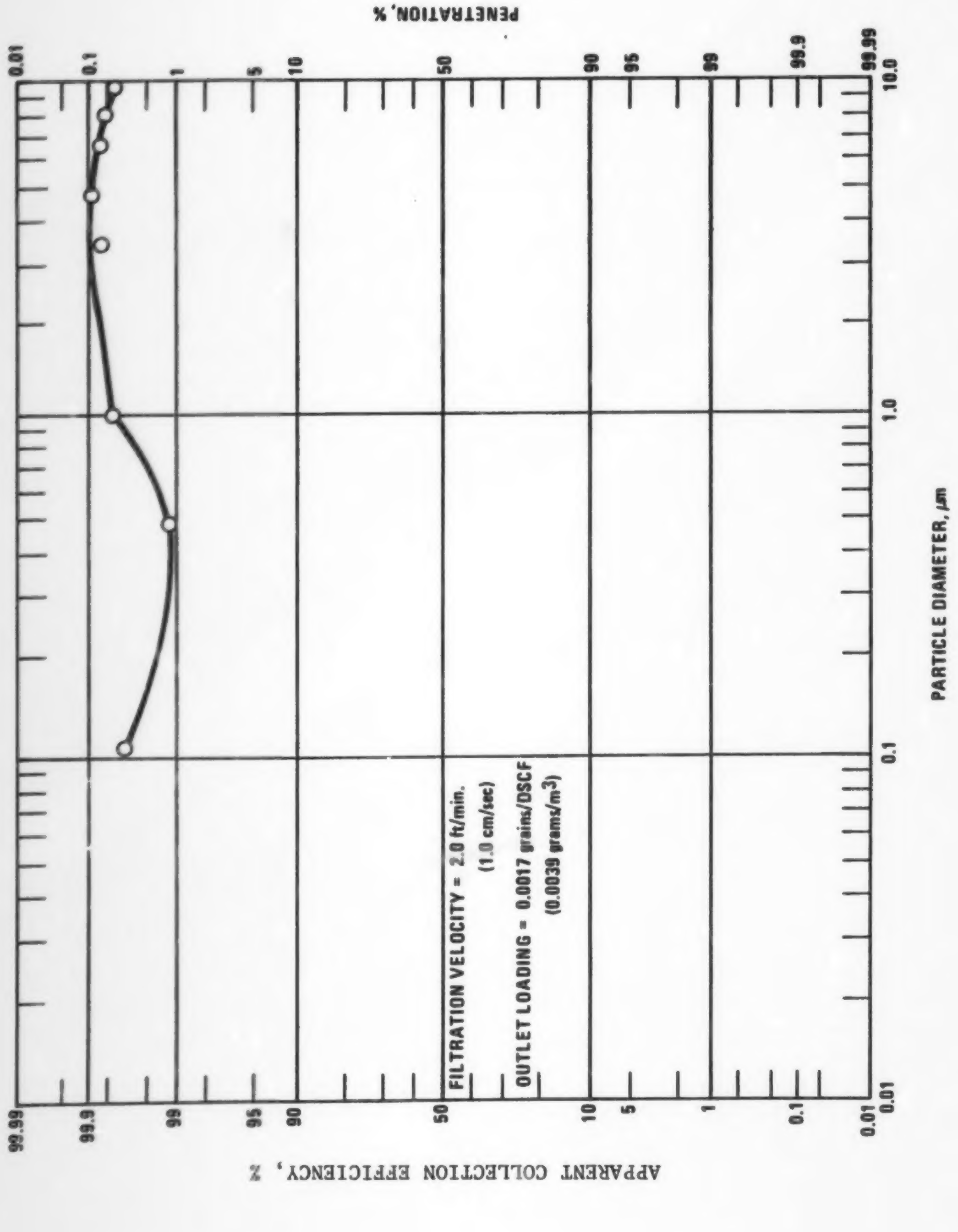


Figure 7 . Baghouse performance at Sunbury Steam Electric Station.

PARTICULATE REMOVAL CHARACTERISTICS OF THE 170 MW MOHAVE SCRUBBERS

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PARTICULATE REMOVAL CHARACTERISTICS
OF THE
170 MW MOHAVE SCRUBBERS

1.0 ABSTRACT

This paper presents the particulate removal data obtained during the 170 MW Test Modules Program at the Mohave Generating Station. Three scrubber configurations are evaluated: Vertical TCA, Vertical PPA, and Horizontal Crossflow. The effects of scrubber variables such as turndown ratio, flue gas pressure drop, liquid to gas contacting ratio and packing characteristics on particulate removal are examined.

A method for predicting particulate removal based on the percentage of electric power consumed by the scrubber and the inlet grain loading characteristics is developed from impaction collection theory and Test Program data. At design operating condition, the Vertical TCA Module required about 35% more electrical power than the Horizontal Module to achieve the same degree of particulate removal.

2.0 INTRODUCTION

Between July, 1971 when pilot plant testing was initiated at the Mohave Generating Station and July, 1975 when the 170 MW Test Modules Program was completed, the participants of the Navajo and Mohave Power Projects funded a scrubber development and evaluation project which has significantly advanced the understanding of scrubber technology. The results of this evaluation have been published,in part, at several EPA Flue Gas Desulfurization Symposium conferences, [1,2,3,4]. The objectives of the 170 MW Test Modules Program were to evaluate the SO_2 and particulate removal performance of more than one full-scale scrubber, as well as to determine the practical feasibility of scrubbers for commercial service, including cost, operating flexibility, maintenance requirements, materials of construction, chemical control problems, and waste disposal factors.

The purpose of this paper is to describe the particulate removal characteristics of the 170 MW Mohave Test Modules. During the test program, a total of 154 inlet/outlet particulate samples were obtained during tests of three different 170 MW scrubber configurations. The scrubbers were located downstream from the cold-side electrostatic precipitators, and were designed to treat 450,000 SCFM of flue gas (equivalent to 170 MW at the Mohave Generating Station).

3.0 TEST MODULE DESCRIPTION

The 170 MW Vertical Module, located on Unit 1, was tested in the Turbulent Contacting Absorber (TCA) configuration and the Polygrid Packed Absorber (PPA) configuration. The TCA configuration, shown in Figure 1, consisted of four stages of thermoplastic rubber balls supported on stainless steel grids at four foot intervals. The balls were contained in compartments, with 15 compartments at each of the four levels. Although various levels of ball depths were tested, the compartments were initially filled to the one-foot level with approximately 1,600,000 balls. The TCA configuration was tested with both three and four stages of balls, and at ball depths of 6, 10, and 12 inches in the four stage configurations. The PPA configuration consisted of two and three stages of fixed "eggcrate" packing supported at four foot intervals. The packing consisted of plastic grids 1¼" thick with 2" square openings, with the grids stacked to a depth of approximately 17 inches in each stage.

The 170 MW Horizontal Module, located on Unit 2, was tested with 1,2,3,4 and 5 spraying stages. The basic four stage configuration, shown in Figure 2, consisted of crossflow spray discharged from a row of 36 externally-mounted nozzles at each stage. The Horizontal Module did not contain packing, but consisted of four stages of cross flow spray. The slurry was cycled through the scrubber in a countercurrent manner. That is, the fresh lime slurry

FIGURE 1

170 MW VERTICAL MODULE (4 STAGE TCA)

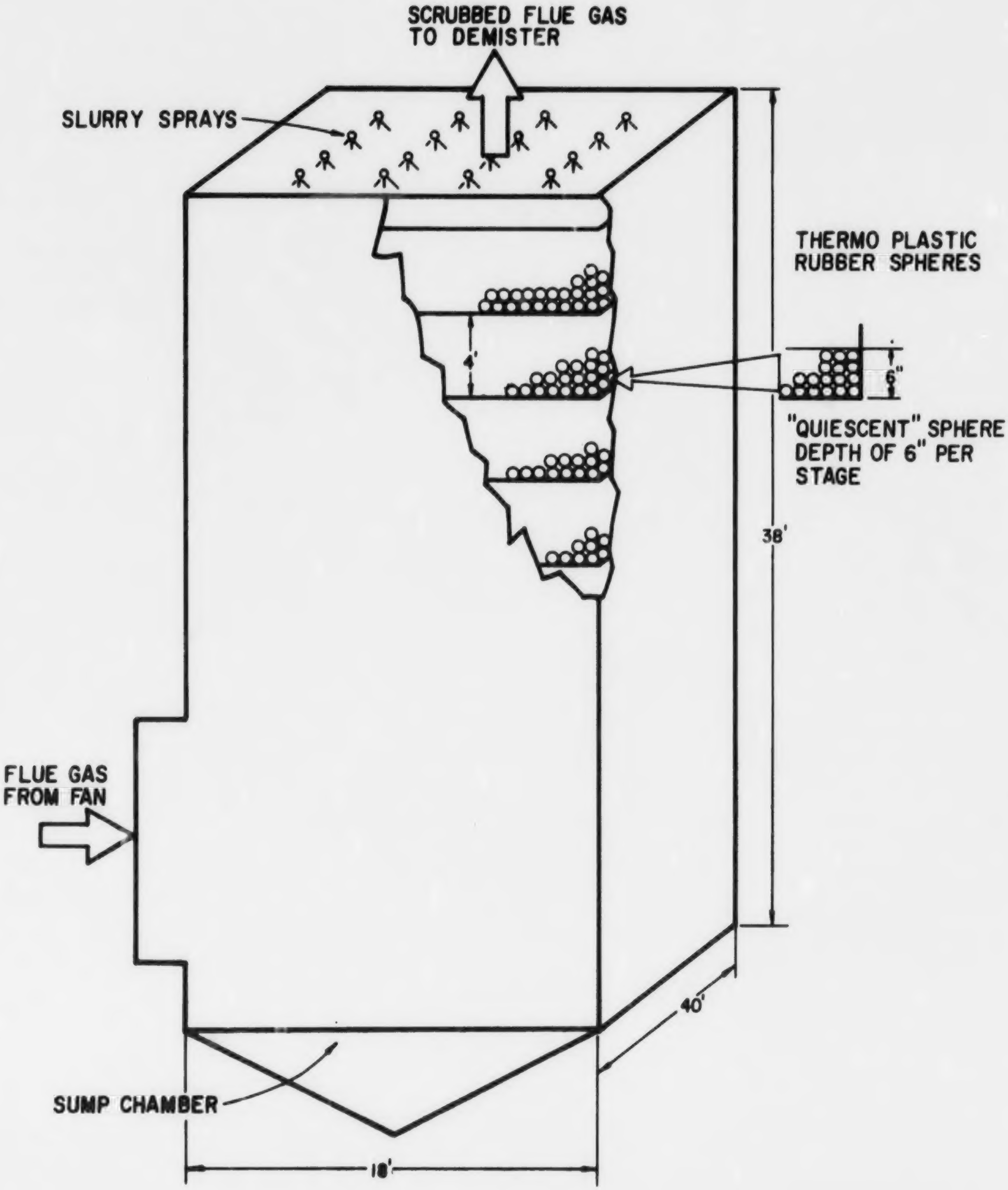
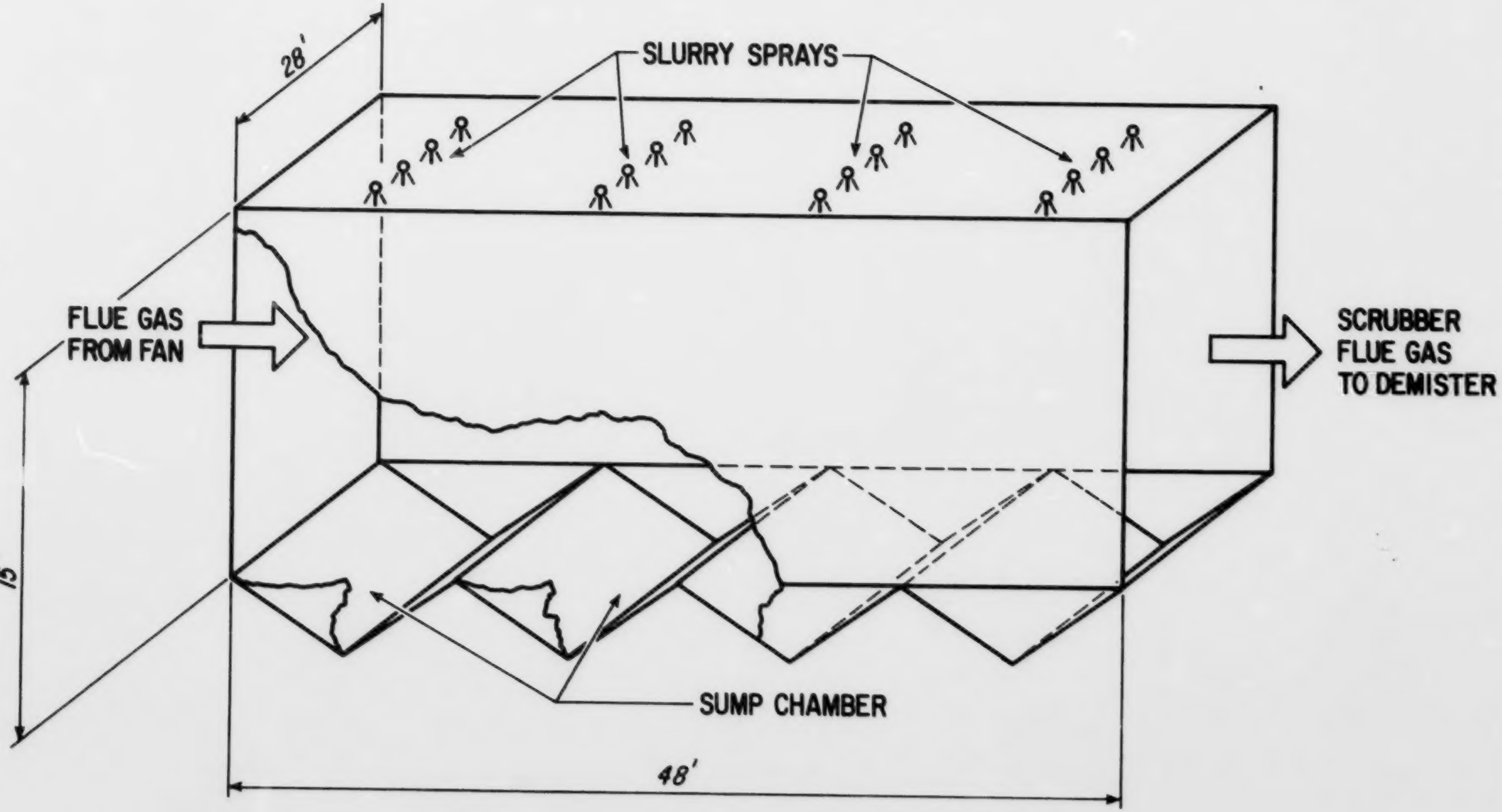


FIGURE 2

170 MW HORIZONTAL MODULE
(4 STAGE)



from the mix tank was first sprayed across the flue gas at the fourth stage, or discharge end of the scrubbing chamber. The same liquid was successively collected and pumped to the third, second, and first stages and successively depleted of alkalinity. By the time the slurry reached the first stage collection hopper for return to the lime mix tank, it was almost completely depleted of any excess alkalinity. The recirculating slurry flow rate was normally 9,000 gpm, but the slurry was mechanically pumped four times per circuit for a total installed pump capacity of 36,000 gpm.

Both the Horizontal and Vertical Modules were designed for identical test conditions, using the same sources of makeup water and removing particulate matter resulting from the combustion of low sulfur western coal from the Black Mesa mine near Kayenta, Arizona.

4.0 OBSERVED SCRUBBER PERFORMANCE

4.1 Particulate Removal at Design Operation Conditions

The observed operating conditions of the three full-scale scrubbers are indicated in Table 1. The outlet gas flowrate was greater than the inlet due to water vapor evaporated inside the scrubber. In addition, the Horizontal Module used an indirect reheater which injected approximately 80,000 SCFM of hot ambient air into the outlet flue gas stream. The pressure drop indicated in Table 1 is the flue gas pressure drop across the scrubbing chamber itself, not including additional pressure drop due to ductwork, mist eliminator and reheater restrictions. The electric power consumption in Table 1 is the total electric power measured for the entire scrubbing system, and includes control room lighting and air conditioning, thickener rake drive, tank agitators, process water pumps and other miscellaneous power requirements. However, these miscellaneous factors are estimated to be about 5% of the total, while fan and slurry pump power requirements accounted for the remaining 95% of the total.

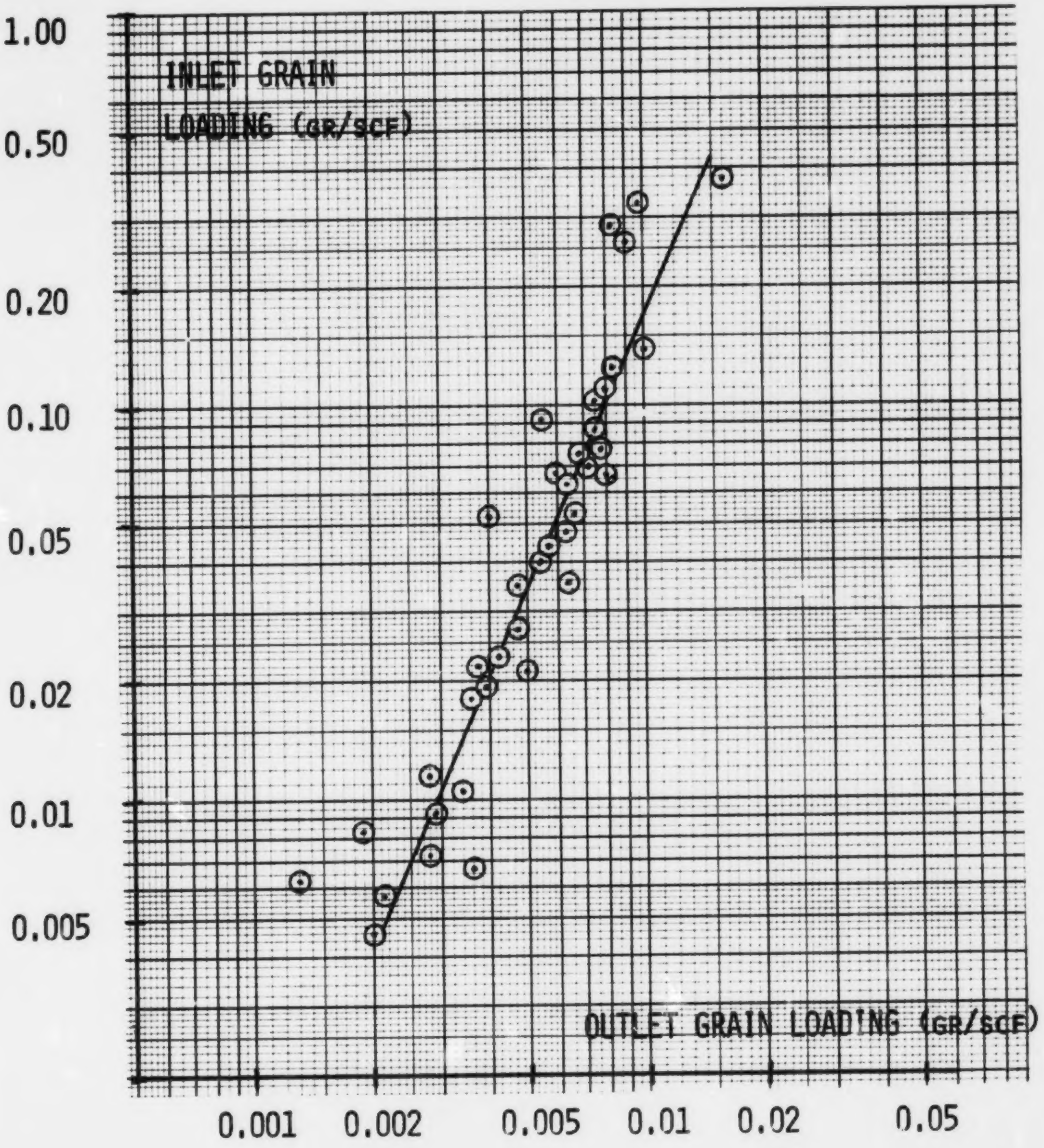
The observed outlet grain loadings obtained at the design operating conditions for the Horizontal, Vertical TCA and Vertical PPA scrubbers are plotted as a function of inlet grain loading in Figures 3, 4 and 5, respectively. At an inlet grain loading of 0.10 gr/SCF, all three scrubber configurations demonstrated a relatively high degree of particulate removal between 90% and 93%. As will be pointed out in the discussion, the percentage particulate removal changed with inlet grain loading due to changes in the average particle size and particle size distribution.

TABLE 1

OPERATING CONDITIONS AT DESIGN GAS FLOWRATE

	HORIZONTAL	VERTICAL TCA	VERTICAL PPA
INLET GAS FLOWRATE	450,000 SCFM	450,000 SCFM	450,000 SCFM
OUTLET GAS FLOWRATE	560,000 SCFM	480,000 SCFM	480,000 SCFM
NUMBER OF STAGES	4	4	3
L/G RATIO (GAL/MSCF)	21 (PER STAGE)	36	60
SCRUBBER GAS VELOCITY	22 FT/SEC	12 FT/SEC	12 FT/SEC
PRESSURE DROP	1.0 IN. H_2O	14.0 IN. H_2O	12.2 IN. H_2O
NOZZLE PRESSURE	35 PSIG	15 PSIG	30 PSIG
POWER CONSUMPTION	2.6 MW	3.4 MW	3.9 MW
TYPE OF REHEAT	INDIRECT	DIRECT	DIRECT

FIGURE 3 INLET vs. OUTLET GRAIN LOADING
170 MW HORIZONTAL MODULE



AVERAGE OPERATING CONDITIONS

INLET FLOWRATE: 450,000 SCFM

OUTLET FLOWRATE: 560,000 SCFM

L/G (PER STAGE): 21 GAL/MSCF

NOZZLE PRESSURE: 35 PSIG

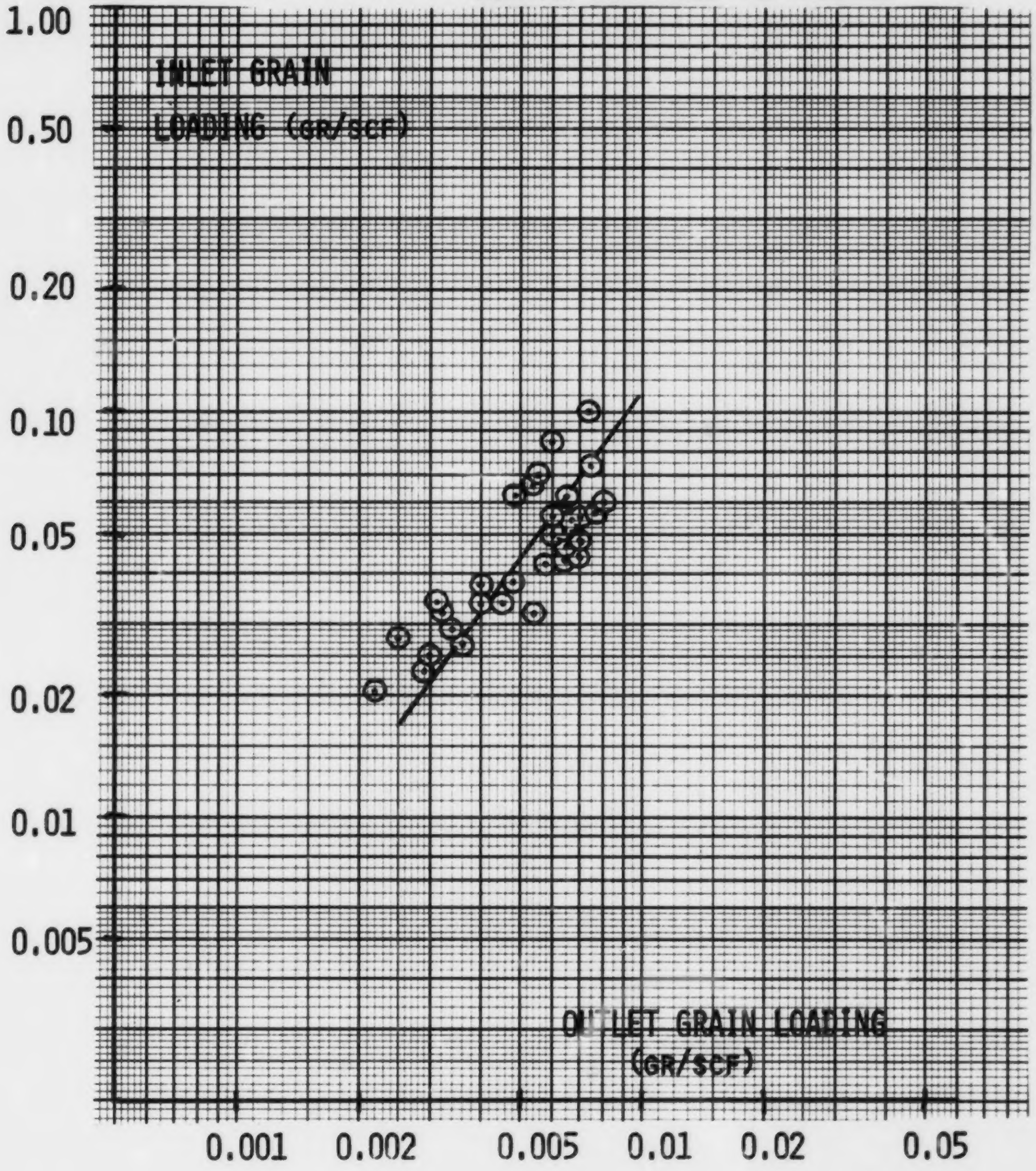
NO. OF STAGES: 4

POWER REQ'D: 2.6 MW

ΔP: 1.0 IN. H_2O

GAS VELOCITY: 22 FPS

FIGURE 4 INLET vs. OUTLET GRAIN LOADING
170 MW VERTICAL TCA MODULE



AVERAGE OPERATING CONDITIONS

INLET FLOWRATE: 450,000 SCFM

OUTLET FLOWRATE: 480,000 SCFM

L/G RATIO: 36 GAL/MSCF

NOZZLE PRESSURE: 15 PSIG

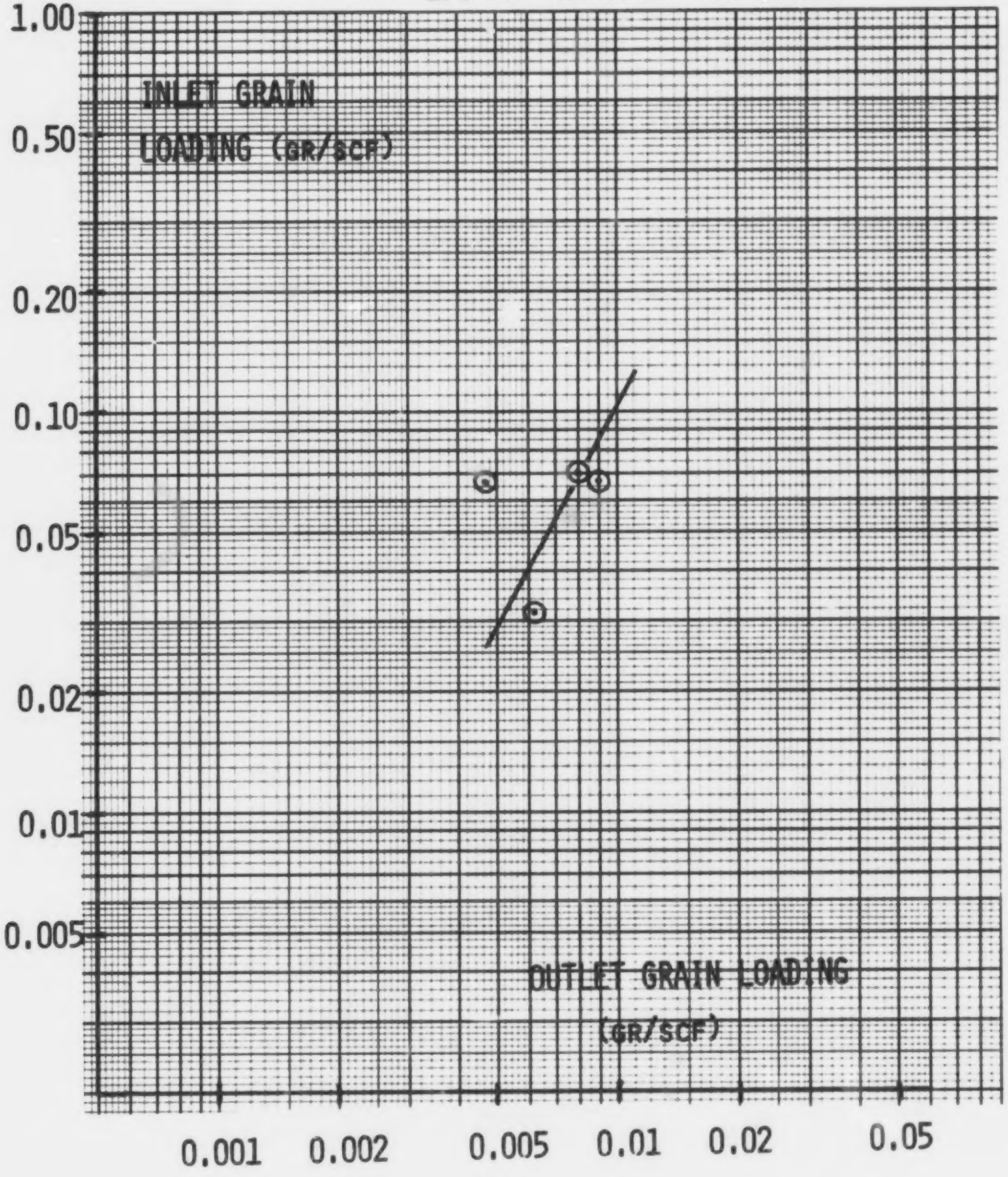
NO. OF STAGES: 4

POWER REQ'D: 3.4 MW

ΔP: 14 IN. H_2O

GAS VELOCITY: 12 FPS

FIGURE 5 INLET vs. OUTLET GRAIN LOADING
170 MW VERTICAL PPA MODULE



AVERAGE OPERATING CONDITIONS

INLET FLOWRATE: 450,000 SCFM
OUTLET FLOWRATE: 480,000 SCFM
L/G RATIO: 60 GAL/MSCF
NOZZLE PRESSURE: 30 PSIG

NO. OF STAGES: 3
POWER REQ'D: 3.9 MW
ΔP: 12.2 IN. H_2O
GAS VELOCITY: 12 FPS

4.2 Effect of Turndown

During the Test Program, it was determined that the easiest method of operating the scrubbers was to maintain a constant liquid flowrate and allow the flue gas flowrate to change under simulated load-following conditions. From an engineering design standpoint, this simplification avoids the severe erosion problems encountered with slurry flow control valves. It was decided to determine the effect on particulate removal of gas flowrate turndown with the liquid flowrates held at the design values shown in Table 1.

The observed variation in percentage particulate removal at a constant inlet grain loading of 0.10 gr/SCF is plotted as a function of the gas flowrate in Figure 6. Note that the particulate removal efficiency of the Horizontal Module increases as a function of turndown ratio.

One explanation for this is the change in energy input per unit of flue gas flowrate which occurs with increasing turndown ratio. The primary method of shattering slurry droplets and thus contacting particulate matter differs between the types of scrubbers tested.

The Vertical TCA and PPA Modules require flue gas pressure drop to increase the tendency to shatter slurry droplets. The particulate removal was thus seen to decrease with decreasing pressure drop. For example, at 450,000 SCFM, the particulate removal was about 90% at the design L/G ratios. Under these conditions the pressure drop was 14 inches of water in the TCA and 12.2 inches in the PPA. When the flue gas flowrate was decreased to one-third the design value, the pressure drop decreased to 3.5 inches of water in the TCA and 2.0 inches in the PPA. In spite of the fact that the L/G ratio was now three times the design value, the particulate removal decreased to about 75%.

The Horizontal Module utilized nozzle pressure to shatter the slurry droplets, and particulate removal was thus seen to increase with decreasing flue gas flowrate. As the flue gas flowrate was turned down from 450,000 SCFM to 150,000 SCFM, the pressure drop across the scrubbing chamber decreased from 1.0 inches to water to 0.10 inches. Since the liquid flowrate was constant, a given amount of flue gas was contacted by three times as many droplets at one-third load as at full load. The particulate removal efficiency was observed to increase from 92.5% at full load to 96.5% at one-third load.

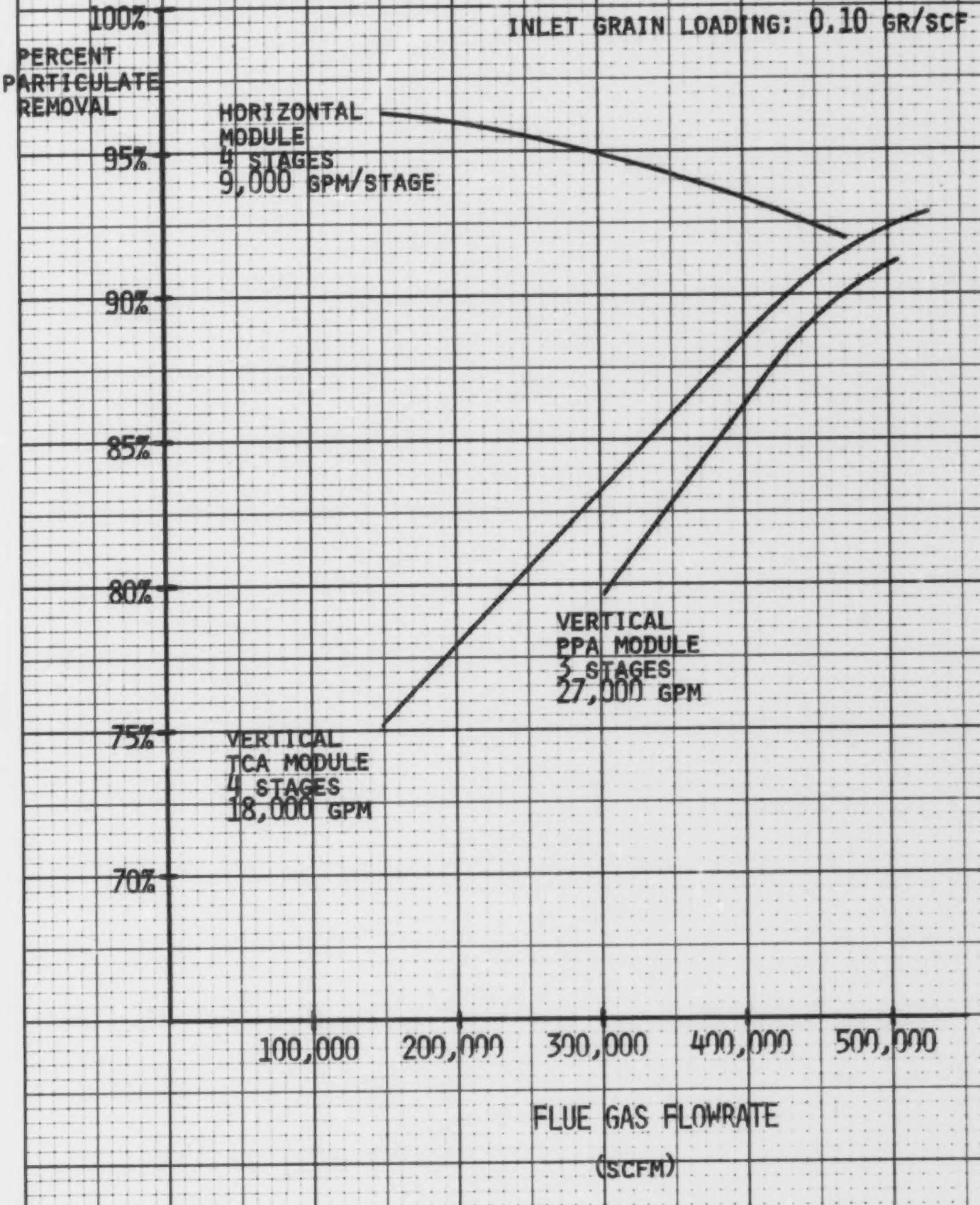


FIGURE 6 EFFECT OF TURNDOWN RATIO ON PARTICULATE REMOVAL

INLET GRAIN LOADING: 0.10 GR/SCF

4.3 Effect of Liquid Flowrate

The effect of increasing the Vertical TCA Module liquid flowrate beyond the levels indicated in Figure 6 was studied as shown in Figure 7. It is seen that an increased liquid flowrate was more effective at a high inlet grain loading than a low inlet grain loading. For example, the same particulate removal achieved with 18,000 gpm at 450,000 SCFM (at an L/G ratio of 40) could be achieved at 350,000 SCFM by doubling the liquid flowrate to 36,000 gpm (at an L/G ratio of 103). The same results could not be obtained at a reduced inlet grain loading of 0.03 gr/SCF.

With the Horizontal Module, particulate removal is only improved as the flue gas flowrate is reduced. It was therefore important to determine how rapidly the particulate removal would decrease with decreasing liquid flowrate. As shown in Figure 8, a reduction in liquid flowrate from 9,000 gpm/stage to 4,500 gpm/stage would result in a decrease in particulate removal from 92% to 78% at an inlet grain loading of 0.10 gr/SCF and from 84% to 66% at an inlet grain loading of 0.03 gr/SCF. These tests were conducted by shutting off individual nozzles to maintain 35 to 40 psig nozzle pressure, and although the spray coverage patterns were undoubtedly altered, the slurry droplet size probably did not change. The required liquid flowrate for a desired degree of particulate removal can thus be estimated from Figure 8.

4.4 Miscellaneous Effects

Several miscellaneous effects were investigated during the Test Program. For example, it was found that increasing the static ball depth from 6" to 12" in the Vertical TCA Module increased the particulate removal.

The effect of variations in nozzle pressure was studied with the Horizontal Module. Particulate removal at an inlet grain loading of 0.032 gr/SCF increased as the nozzle pressure was increased from 15 psig to 30 psig.

Finally, it was confirmed that increasing the number of contacting stages improves the particulate removal efficiency. For example, the particulate removal with the Vertical PPA Module increased as the number of packed stages was increased from two to three. Similar results were obtained with the Horizontal Module, where the effect of four versus five stages was studied in detail. A comparison of fractional collection efficiencies with four and five stages of scrubbing indicated that the extra stage increased the collection of 1.5 micron (aerodynamic) particles. An interesting description of particulate collection in spray tower applications due to condensation effects can be found in Calvert, et al.[5] This may be a partial explanation for the degree of particulate removal achieved with the Horizontal Module.

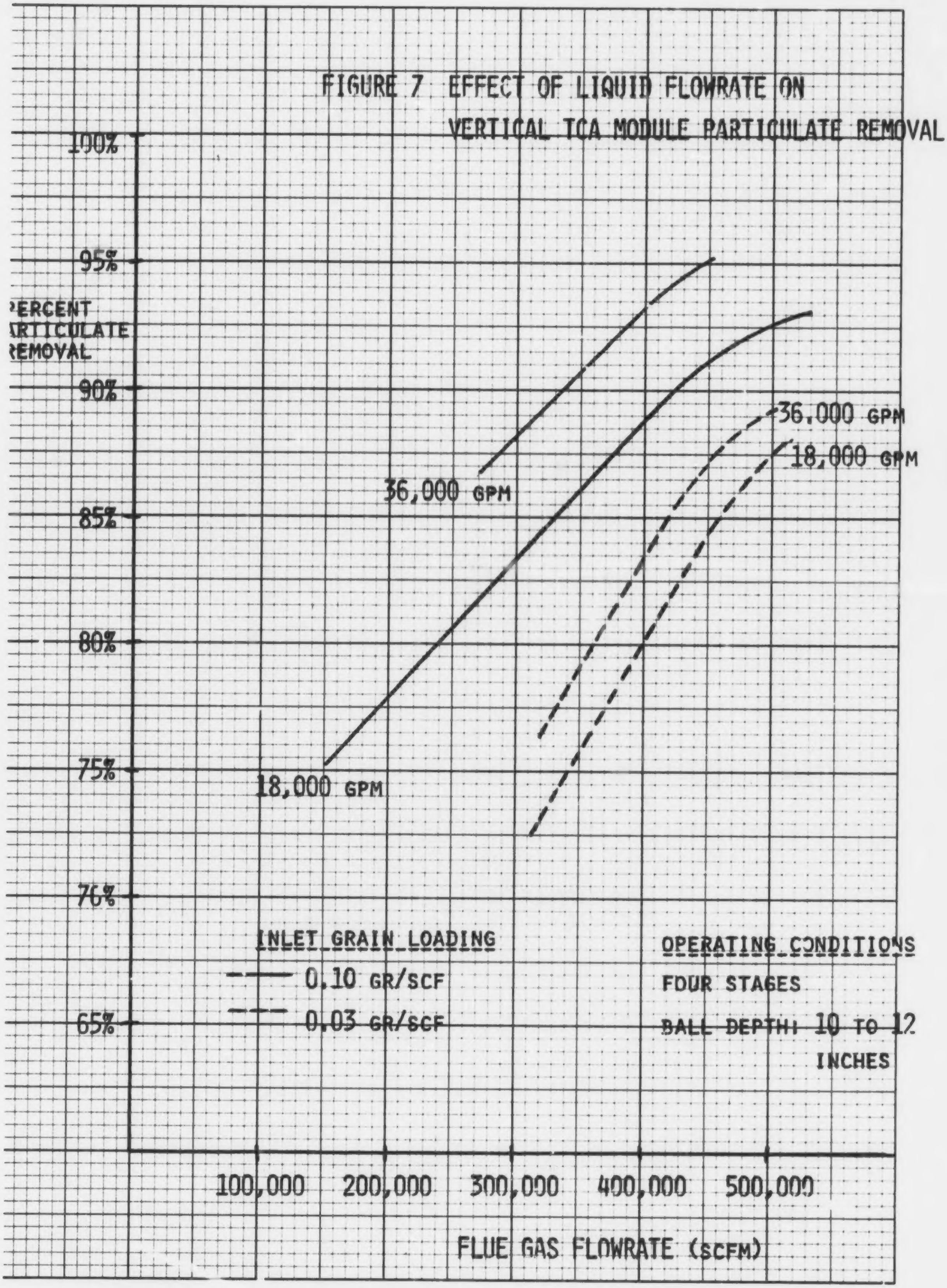


FIGURE 7 EFFECT OF LIQUID FLOWRATE ON VERTICAL TCA MODULE PARTICULATE REMOVAL

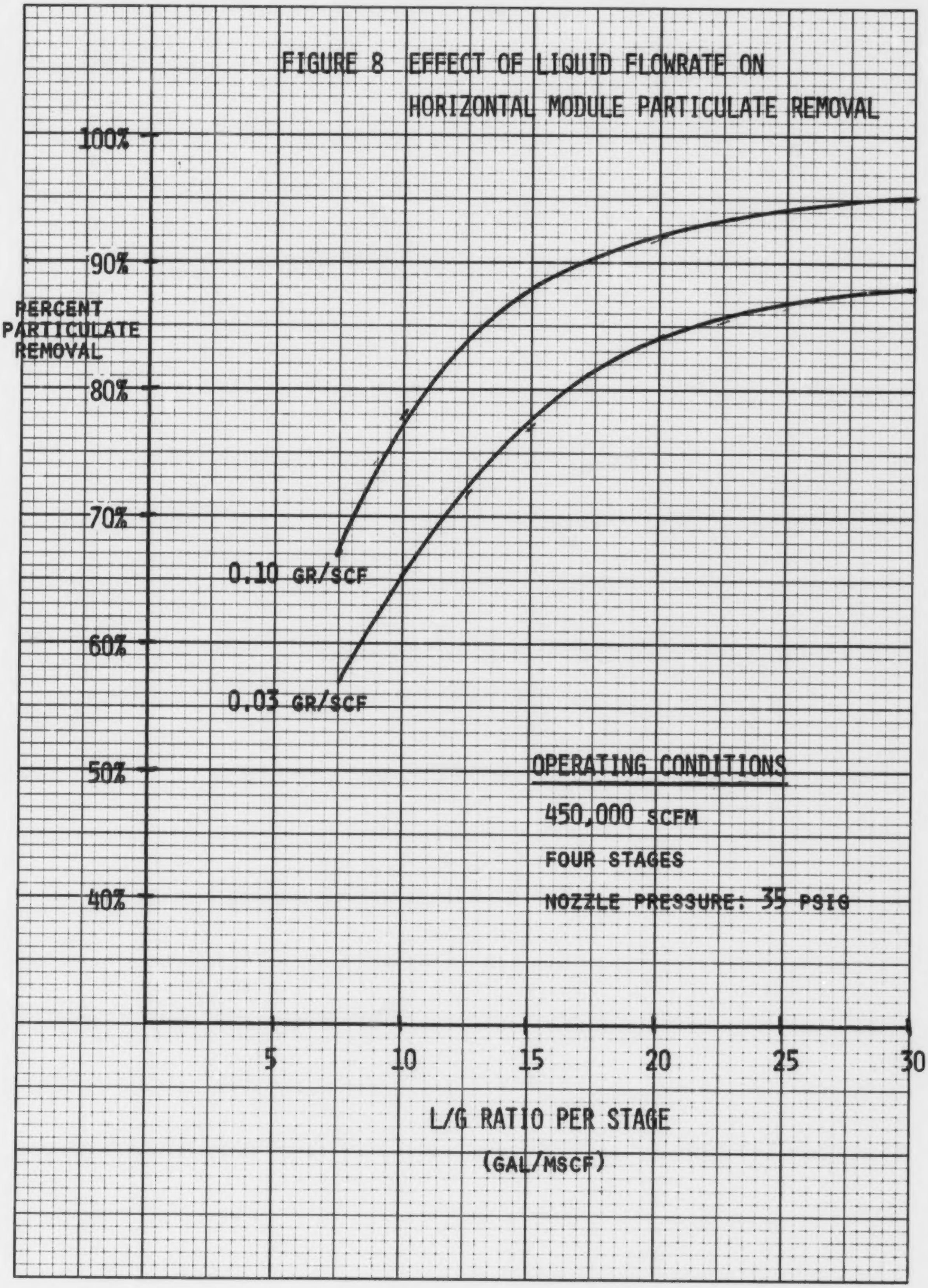


FIGURE 8 EFFECT OF LIQUID FLOWRATE ON HORIZONTAL MODULE PARTICULATE REMOVAL

5.0 DISCUSSION

5.1 Data Acquisition and Reduction

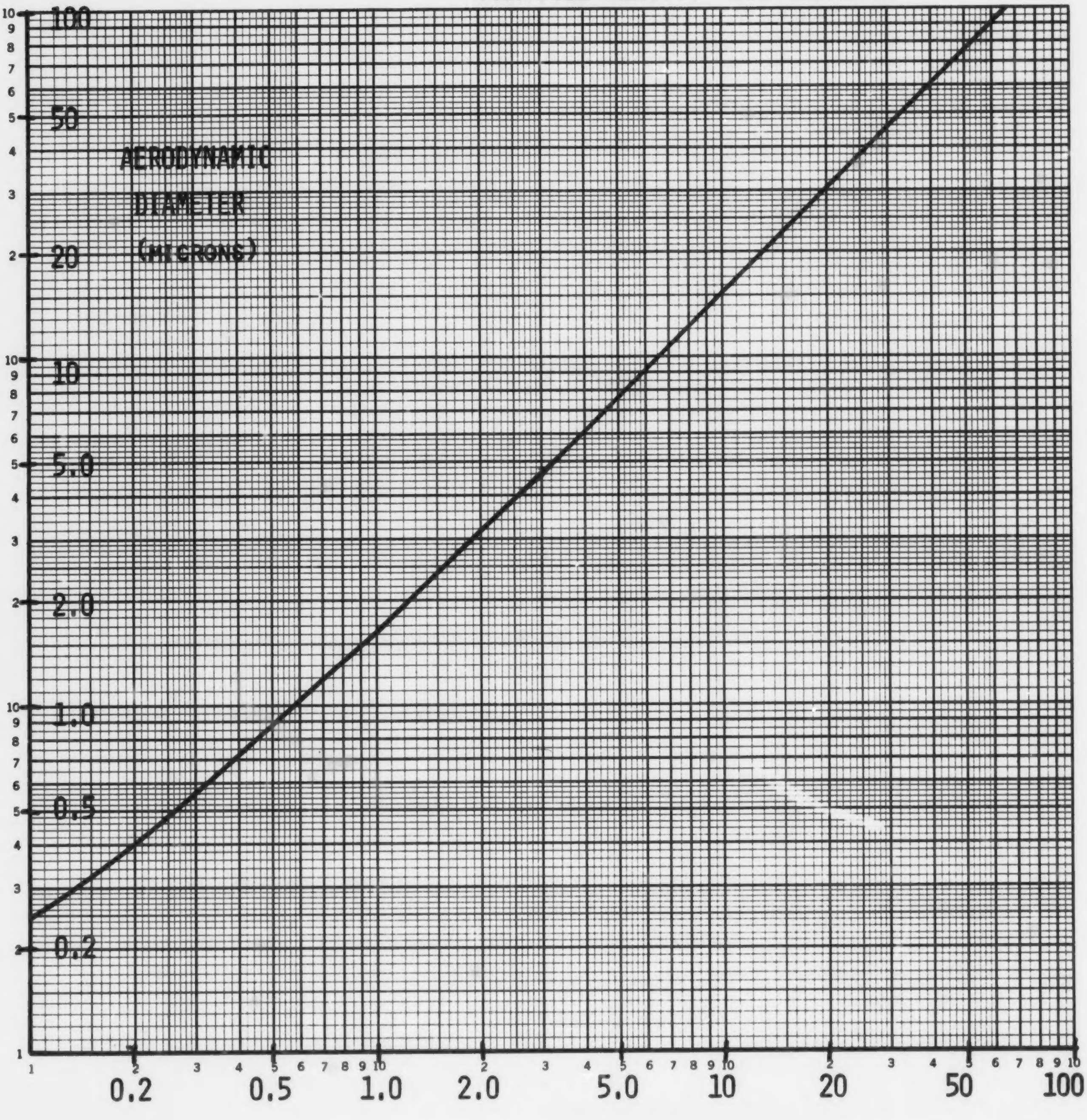
It must be pointed out that getting good particulate data by sampling in large ductwork with cross-sections of 200 square feet or more is not easy. Gas sampling crews of five persons working under ideal conditions were able to obtain one inlet and one outlet isokinetic particulate sample by probe traverse in one day. All testing was subject to unit outages and weather conditions ranging from 70 mph winds in the winter to 130° F temperatures in the summer. All testing was conducted by probe traverse, and electrical discharge from the metal probes to ground were observed at the inlet duct due to electrical charge carried on the inlet flyash particles.

Grain loading measurements were conducted using an alundum thimble with a backup millipore filter, and gas volumes were metered and corrected for water vapor condensation in ice bath condensers. Particle size distribution measurements were obtained with both Anderson and University of Washington cascade impactors. Most of the particle size distribution measurements on the Vertical TCA Module were conducted with Anderson impactors without backup filters while similar measurements on the Horizontal Module were conducted primarily with University of Washington impactors with backup filters. It was found that physically identical impactors do not produce the same results from the same sample and must be individually calibrated using particles of a known size and instrumentation to measure the concentration of particles upstream and downstream from the impactor.

Because of these differences, the particle penetration data for the Horizontal and Vertical TCA Modules at design operating conditions is difficult to compare directly. However, it can be stated with a reasonable degree of confidence that both the Horizontal and Vertical TCA Modules achieved approximately 65% collection of the 1.5 microns (aerodynamic) particles at design operating conditions. An examination of the slopes of figures 3 and 4 would also suggest that the Horizontal Module at design conditions achieved a higher degree of collection efficiency for particles larger than 1.5 microns (aerodynamic) than the Vertical Module at design operating conditions.

The particle size data presented in this paper refers to aerodynamic particle diameter as measured by a calibrated cascade impactors. The aerodynamic particle diameter takes into account particle density and shape factor, and should not be confused with the actual particle diameter. A curve showing the relationship between aerodynamic and actual particle diameter for Mohave Generating Station flyash having a density of 2.2 g/cm^3 and consisting primarily of spherical particles is presented in Figure 9. The term geometric standard deviation is used in the text to identify the distribution of particle sizes about an average size. For a log-normal particle size distribution, a geometric standard

FIGURE 9 AERODYNAMIC vs. ACTUAL PARTICLE DIAMETER
MOHAVE GENERATING STATION FLYASH
PARTICLE DENSITY: 2.2 GRAMS/CM^3



ACTUAL DIAMETER
(MICRONS)

deviation of two means that the size below which 84 weight percent of the particles are found is twice the average size. Thus, for a geometric standard deviation of one, all the particles are the same size. A brief description of particle size distribution analysis is included in Appendix 1 for reference.

As another prerequisite to obtaining high quality, accurate data, detailed grain loading and size distribution studies were conducted to determine the effects of stratification in the ductwork. It was found that below a grain loading of about 0.05 gr/SCF, a variation of only ± 15% in geometric mean particle size could be expected due to stratification at various points in the duct.

5.2 Characteristics of Mohave Flyash

The geometric mean particle size and standard deviation of the flyash at the scrubber inlet changed as a function of grain loading, as shown in Table 2.

An examination of Table 2 indicates that an increase in total grain loading by a factor of twenty will cause an increase in the grain loading of small particles less than 2 microns (aerodynamic) diameter by a factor of only four. This means that changes in grain loading are almost entirely determined by changes in the weight percentage of very large particles. It would thus be expected that the outlet grain loadings from the Test Modules would change over a narrow range for a wide range of changes in inlet grain loading. This has already been shown in Figures 3, 4 and 5.

5.3 Performance Predicted by Inertial Impaction

If interest is restricted to particle diameters more than about 1.0 micron (aerodynamic), then Ranz and Wong[6] have shown that collection by inertial impaction is by far the most important collection mechanism. Particle penetration for many types of control equipment can be expressed as

$$P = \exp\left(-K d_p^{B}\right) \qquad \text{where K is a constant} \qquad (1)$$

Calvert [7,8] has indicated that the exponent B equals 2 for most types of packed towers, sieve plate columns and venturi scrubbers. If the particle penetration is integrated over the entire size distribution, the average penetration can be determined

$$PAV = \int_0^W P \frac{(dw)}{(w)} = \int_0^W \exp\left(-K d_p^{2}\right) \frac{(dw)}{(w)} \qquad (2)$$

where w is the particle weight

TABLE 2

ESTIMATED FLYASH CHARACTERISTICS
MOHAVE GENERATING STATION

SCRUBBER INLET GRAIN LOADING (GR/SCF)	AVERAGE PARTICLE DIAMETER (AERODYNAMIC MICRONS)	GEOMETRIC STANDARD DEVIATION σ_G
0.02	3.0	2.7
0.04	4.0	2.8
0.10	6.0	3.0
0.40	16.0	4.5

Following Calvert, equation (2) can be solved for a log-normal particle size distribution as a function of the inlet size distribution parameters and the so-called "cut diameter". The "cut diameter," dp50, is the particle diameter for which the collection efficiency of the control hardware is 50%. Therefore, more than half of the particles larger than the "cut diameter" will be collected, and conversely, less than half of the particles smaller than the "cut diameter" will be collected. The "cut diameter" characterizes the overall particulate removal efficiency of a control device for known inlet particulate characteristics. The solution to equation (2) is plotted in Figure 10, and gives overall penetration as a function of the inlet particle size distribution characteristics and the "cut diameter" of the particulate control device.

A method for determining the "cut diameter" of the Horizontal and Vertical TCA Modules is given in Figure 11, where "cut diameter" is plotted as a function of the percent of net electrical power required to operate the scrubber. Figure 11 was prepared from the data in Figures 7 and 8 as well as control room measurements of electric power requirements over a wide range of operating conditions. Since the "cut diameter" of a given scrubber depends on the energy supplied for droplet shattering and liquid/gas contacting, it is reasonable to expect that the "cut diameter" would correlate with percent of net electrical power required to operate the scrubber. A control device using less electrical power to achieve the same "cut diameter" would be a more efficient device in terms of particulate removal. The data in Figure 11 includes penalties for inefficient power consumption, such as pressure drop losses due to ductwork restrictions, pumping losses due to elevation differences and less than 100% mechanical efficiency of rotating equipment.

As shown in Figure 11, a "cut diameter" of a 1.0 microns (aerodynamic) could be achieved at 450,000 SCFM for an electric power consumption factor of 1.7% with the Horizontal Module and 2.3% with the Vertical TCA Module. Both scrubbers could achieve this result, but one is 35% less efficient than the other. When turndown with the Vertical TCA Module is taken into account, the electrical power consumption resulting from an increased liquid flowrate required to achieve the same "cut diameter" increases significantly.

For the engineer, this method of predicting overall particulate removal knowing the inlet particulate characteristics, the type of scrubber and its electric power consumption characteristics can be a useful design tool.

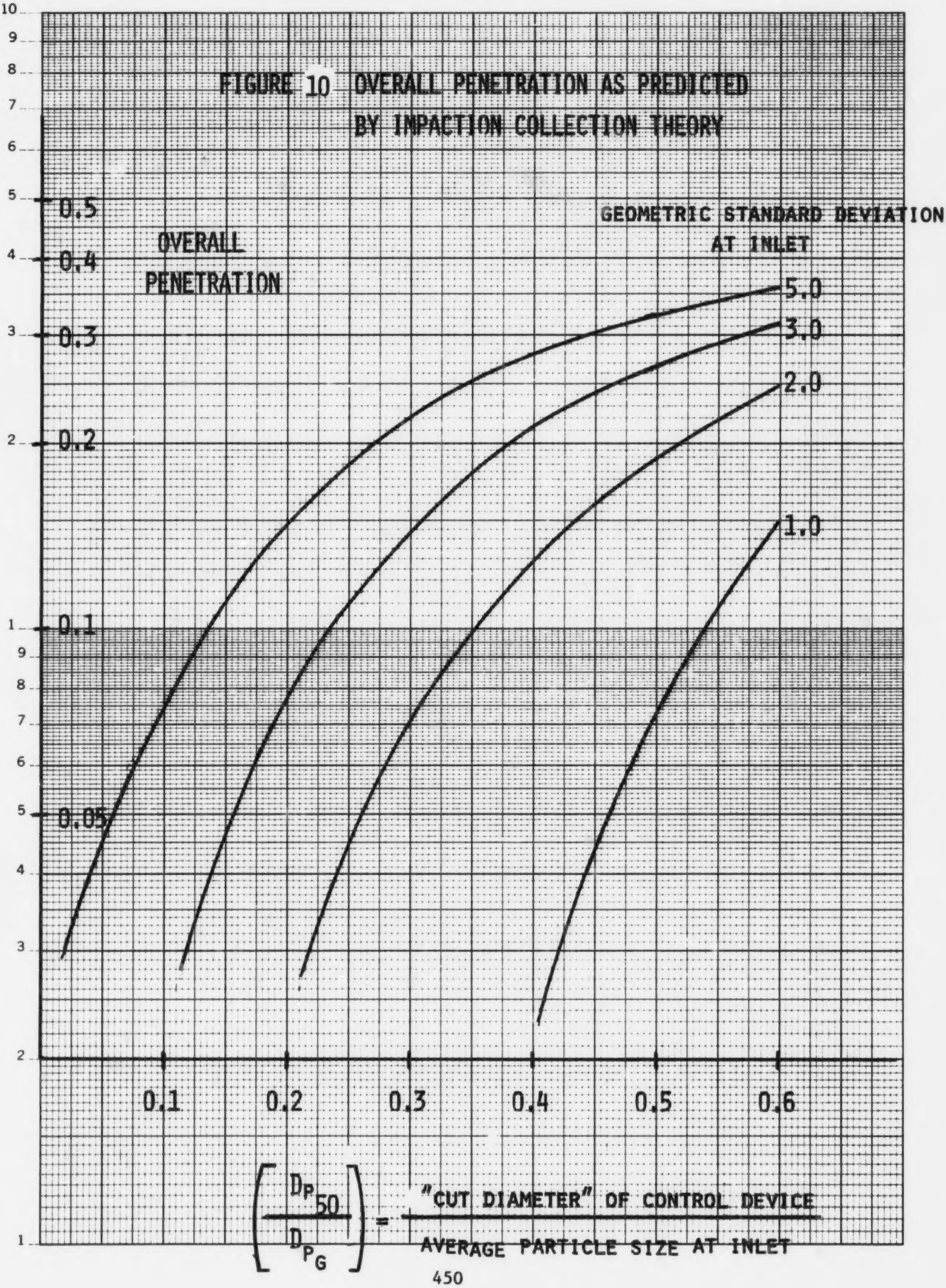


FIGURE 10 OVERALL PENETRATION AS PREDICTED BY IMPACTION COLLECTION THEORY

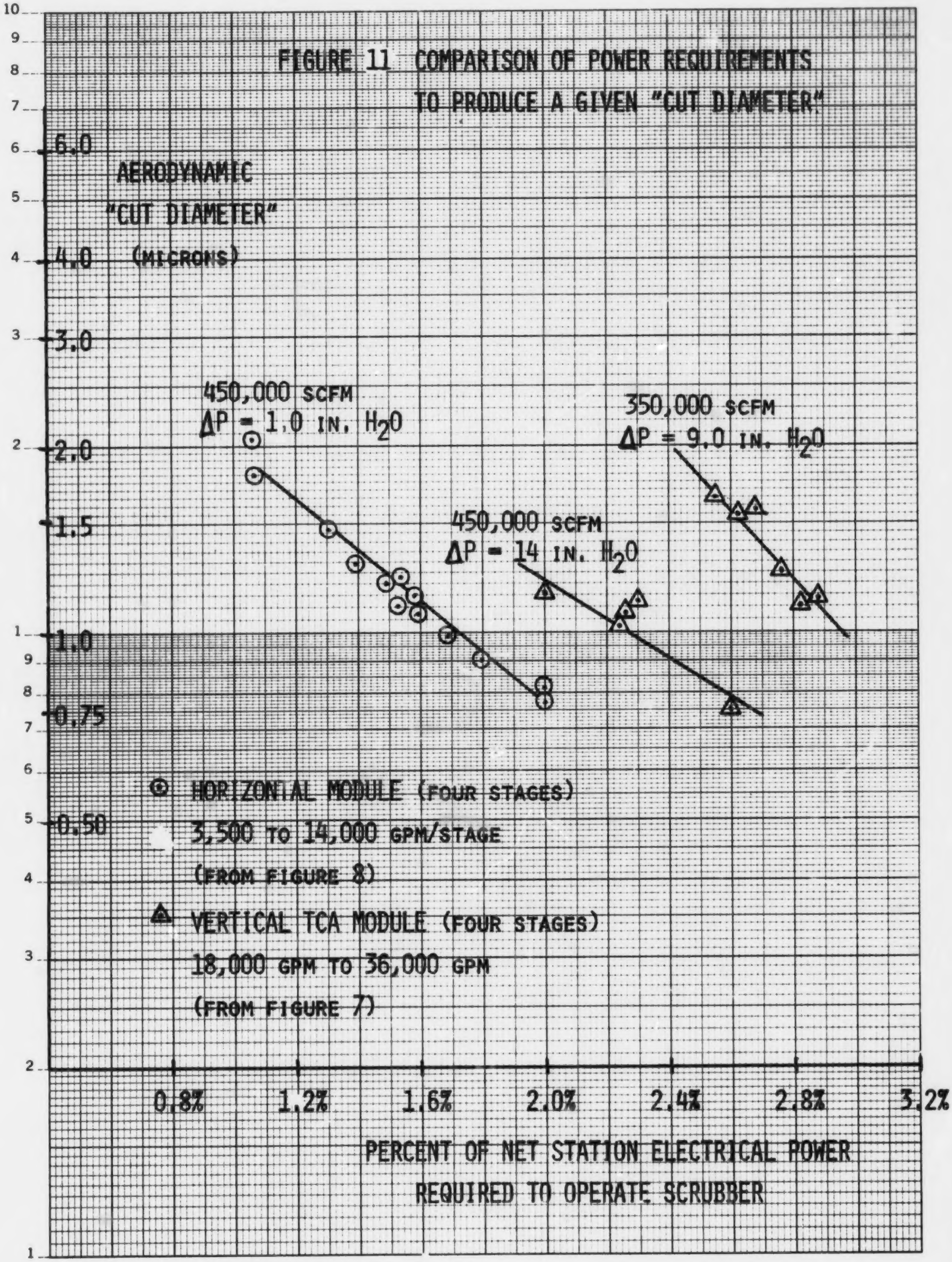


FIGURE 11 COMPARISON OF POWER REQUIREMENTS TO PRODUCE A GIVEN "CUT DIAMETER"

Example 1

Horizontal Module at 2.6 MW and 450,000 SCFM
Estimated MW capacity at 450,000 SCFM: 170 MW
Power consumption factor = 1.53%
"Cut diameter" from Figure 11: dp50 = 1.20 microns
Select inlet grain loading of 0.04 gr/SCF
Inlet particulate characteristics from Table 2

$$dpg = 4.0 \text{ microns and } \sigma g = 2.8$$

$$\frac{(dp50)}{(dpg\ )} = 1.20/4.0 = 0.30$$

Overall penetration from Figure 10: PAV = 0.125
Predicted Particulate Removal: 87.5%
Check from Figure 3 gives 86.5%

Example 2

Vertical TCA Module at ΔP = 9.0 in H_2O, liquid flowrate of 10,000 gpm and gas flowrate of 350,000 SCFM
Estimated Module Power = 3.2 MW
Estimated MW capacity at 350,000 SCFM = 130 MW
Power consumption factor = 2.46%
"Cut diameter" from Figure 11: dp50 = 1.85 microns
Select inlet grain loading of 0.04 gr/SCF
Inlet particulate characteristics from Table 2

$$dpg = 4.0 \text{ microns and } \sigma g = 2.8$$

$$\frac{(dpg50)}{(\ dpg\ )} = 1.85/4.0 = 0.46$$

Overall penetration from Figure 10: 0.23
Predicted particulate removal: 77%
Check from Figure 7 gives approximately 76% for an inlet grain loading of 0.03 gr/SCF

6.0 CONCLUSIONS

Several conclusions can be drawn from the data presented in this paper:

1. For the 170 MW scrubbers tested with Mohave Generating Station flyash, more than 90% total particulate removal can be achieved with mobile bed, fixed bed and spray chamber scrubbers operating under design conditions at 450,000 SCFM and an inlet grain loading of 0.1 gr/SCF.

2. The effect of turning down the flue gas flowrate without changing the liquid flowrate increases the particulate removal of the Horizontal Module and significantly decreases the particulate removal of the Vertical TCA and PPA Modules. This could limit the particulate removal effectiveness of the Vertical Module configurations during normal load following operation.

3. The effect of changes in liquid flowrate on particulate removal is more pronounced with the Horizontal Module than the Vertical TCA Module.

4. Of all the miscellaneous effects tested with the Vertical Module, including stages of packing in the PPA configuration and static ball depth in the TCA configuration, only minor changes in particulate removal were observed.

5. Increases in the number of stages and the nozzle pressure with the Horizontal Module cause a significant increase in particulate removal. However, incremental improvements in particulate removal performance with more stages or higher nozzle pressures exhibit practical limits according to the law of diminishing returns.

6. Large increases in grain loading at the scrubber inlet causes relatively small increases in grain loading at the scrubber outlet. This is due to high collection efficiency for large particles. For example, at design operating conditions the Horizontal and Vertical TCA Modules both achieve more than 95% removal for particles larger than 6.0 microns (aerodynamic).

7. At design operating conditions, the Vertical TCA Module requires about 35% more electrical power than the Horizontal Module to achieve the same degree of particulate removal.

7.0 ACKNOWLEDGMENT

The Test Modules Program was a joint venture of the Navajo and Mohave Power Project participants who are listed below:

Salt River Project Agriculture Improvement and Power District

Arizona Public Service Company

Department of Water and Power of the City of Los Angeles

Nevada Power Company

Tucson Gas and Electric Company

Bureau of Reclamation of the U.S. Department of the Interior

Southern California Edison Company

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The conclusions presented in this paper represent the personal opinions of the authors and are not intended to represent the opinions or position of any of the project participants.

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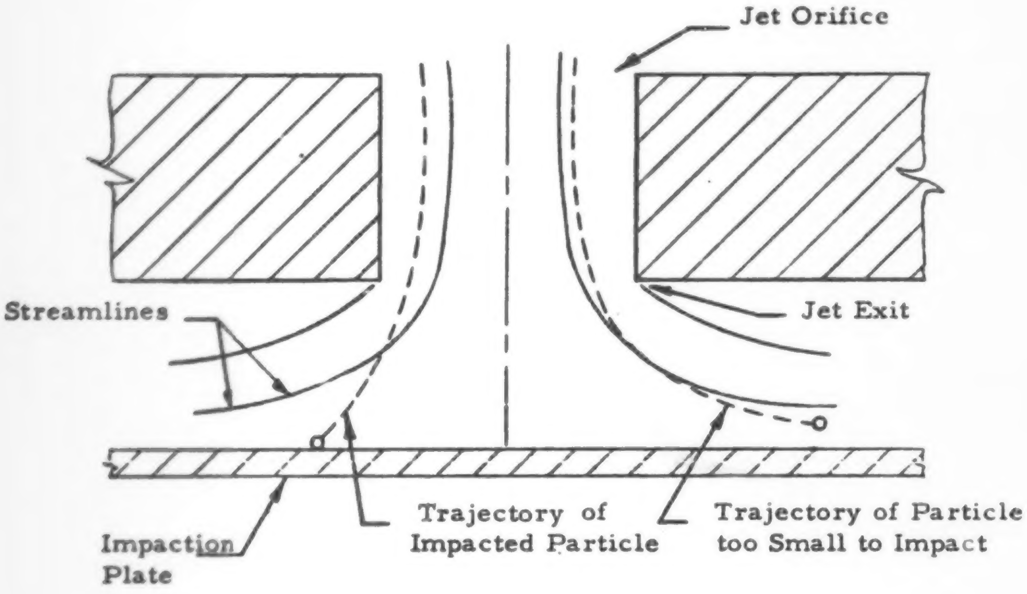
9.0 APPENDIX 1 - PARTICLE SIZE DISTRIBUTION ANALYSIS

Description of Cascade Impactors

Reviewing the theory which underlies how a cascade impactor works, a typical device is a series of plates containing holes of a given size on each plate. The holes act as gas jets when a sample is drawn through the cascade, and the gas jet from the holes in an upstream plate, or stage, impinge on areas which do not contain holes on the downstream stage.

The efficiency of particle separation increases as the particle diameter and gas velocity increase and as jet diameter decreases. Thus one can set up a series (or cascade) of impinging jets with increasing collection efficiency so that smaller particles would be collected in succeeding stages of the cascade. While the separating mechanism is different, the effect is just the same as using a series of screens with decreasing opening sizes. One arrangement of a series of perforated plate and impacting surface pairs as used in a typical impactor has the collection efficiency characteristics shown in Figure 9.1.


Jet Orifice
Streamlines
Jet Exit
Trajectory of Impacted Particle
Trajectory of Particle too Small to Impact
Impaction Plate

SCHEMATIC OF AN IMPACTOR STAGE

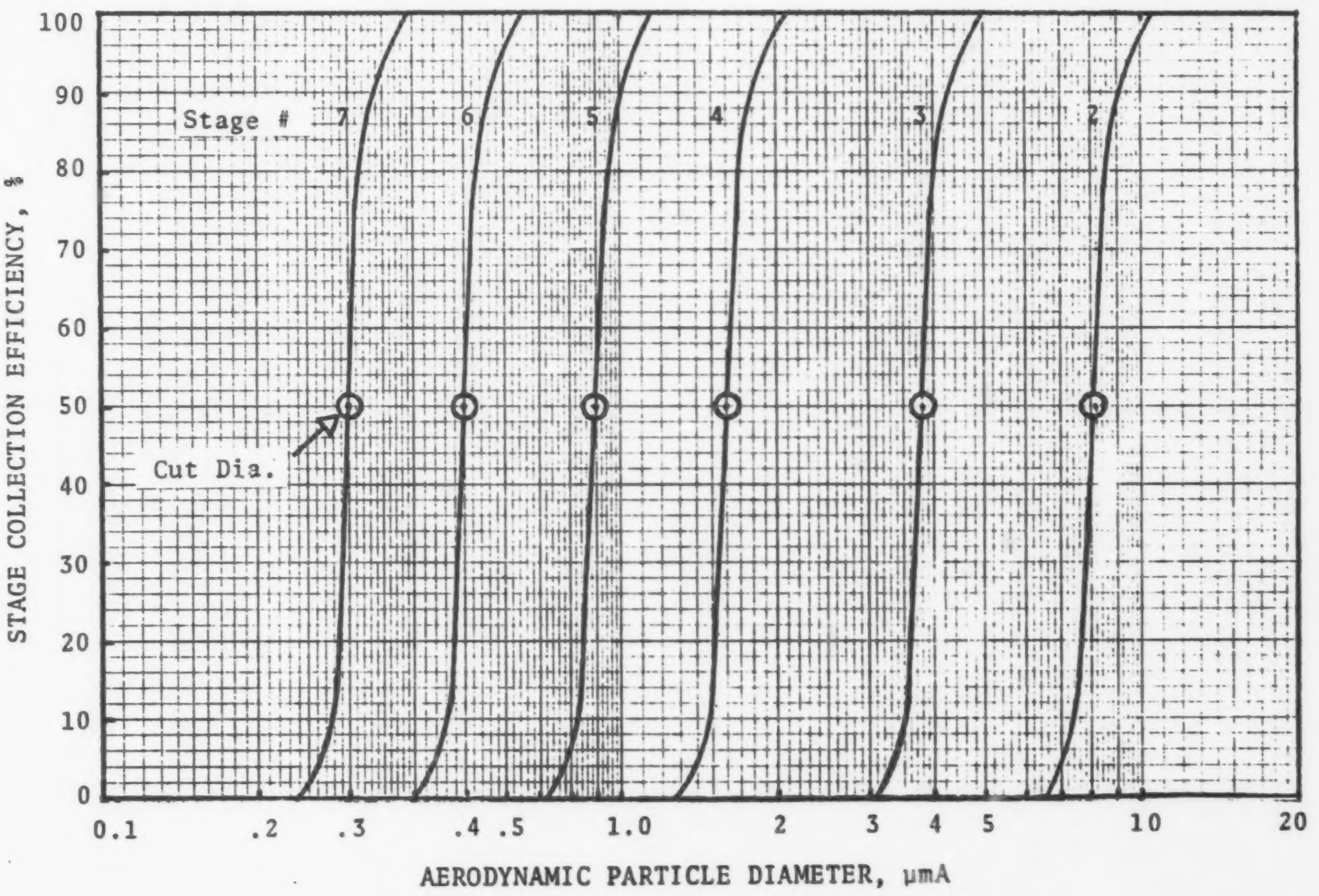


Figure 9.1 Cascade Impactor Stage Efficiencies (for 1 CFM and standard conditions).

Note that the particle diameter used is called "aerodynamic diameter" and it has the units of "aerodynamic microns", µmA. This is the effective diameter for particle separation by inertial impaction and it takes into account the effects of particle density and particle "slip" between gas molecules. The collection efficiency of an impinging jet depends on the inertial impaction parameter, K_p, and the distance the jet travels from the orifice to the impaction plate. For ratios of jet length to diameter between 1 and 4 there is little difference in collection efficiency. In that range of jet lengths the experimentally determined cut diameter is given by:

$$d_{p50} = \left(K_{p50}\right)^{\frac{1}{2}} \left(\frac{9ud_h}{\rho_p C' u_h}\right)^{\frac{1}{2}} \times 10^4 \qquad (1)$$

and $$\left(K_{p50}\right)^{\frac{1}{2}} = 0.45 \qquad (2)$$

Where:

d_{p50} = diameter of particle collected with 50% efficiency µm

K_{p50} = inertial impaction parameter for 50% efficiency collection of d_{p50}

ρ_p = particle density, g/cm^3

C' = Cunningham "slip" correction factor

u_h = jet velocity, cm/sec

u = gas viscosity, Poise

d_h = jet diameter, cm

All of the particle properties can be lumped together in the aerodynamic particle diameter, which is defined as:

$$d_{pa} = d_p\left(\rho_p C'\right)^{\frac{1}{2}} \qquad (3)$$

The aerodynamic cut diameter, d_{pc}, is therefore given by:

$$d_{pc} = \left(K_{p50}\right)^{\frac{1}{2}} \left(\frac{9ud_h}{u_h}\right)^{\frac{1}{2}} \times 10^4 = 0.45 \left(\frac{9ud_h}{u_h}\right)^{\frac{1}{2}} \times 10^4 \qquad (4)$$

If collection efficiency for round jet impaction is plotted against the ratio of particle aerodynamic diameter to "aerodynamic cut diameter", one curve describes all cases within the usual range of jet length to diameter ratios. The cut diameter is the particle size which is collected at 50% efficiency. Figure 9.2 is such a plot and it describes the shape of each of the individual curves in Figure 1. The sharpness of fractionation possible with an impactor is clearly shown in Figure 9.2. Particle collection efficiencies for diameters other than the cut size can be found by the use of Figure 9.2. This relationship as well as the value of the impaction parameter for the cut size (equation (2)) has been determined experimentally.

Effects of Temperature and Pressure

While the effect of gas flow rate is accounted for by equation (1), temperature and pressure have effects whose prediction requires additional information. Temperature affects both the gas viscosity and the Cunningham factor, C'. Pressure has a significant effect on C' but not much on viscosity. Gas composition has an effect on both viscosity and C' but these variations are not large in the usual air and flue gases encountered.

The situations one may encounter are as follows:

1. Aerodynamic cut diameter as defined at impactor conditions is affected through effects on gas viscosity and velocity.
2. Actual particle cut diameter, d_p, is computed from aerodynamic cut diameter with compensation for particle density, ρ_p, and C' at impactor conditions.
3. Aerodynamic cut diameter at conditions other than those in the impactor can be computed from the actual diameter with compensation for particle density at C' at the new conditions.

Thus, in order to take temperature, pressure and gas composition effects into account, it is necessary to have additional information relating these parameters to gas viscosity and to the Cunningham factor. Gas viscosity data can be found in standard chemical engineering and chemistry handbooks. The Cunningham correction factor is a complex function of the mean free path of the gas and the particle diameter. It increases as temperature increases and as pressure decreases. For air at standard temperature and pressure it is given approximately by:

$$C' = 1 + \frac{0.165}{d_p} \qquad (5)$$

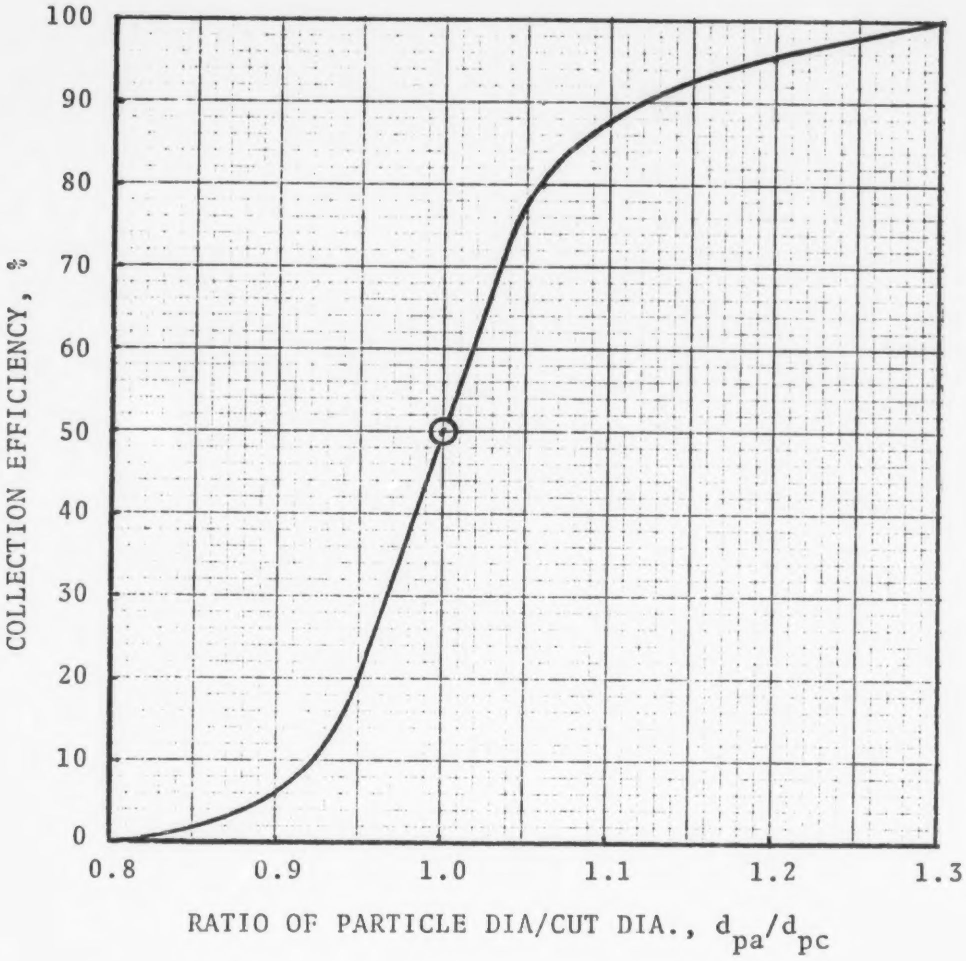


Figure 9.2. Generalized Collection Efficiency - Size Relationship

Illustration

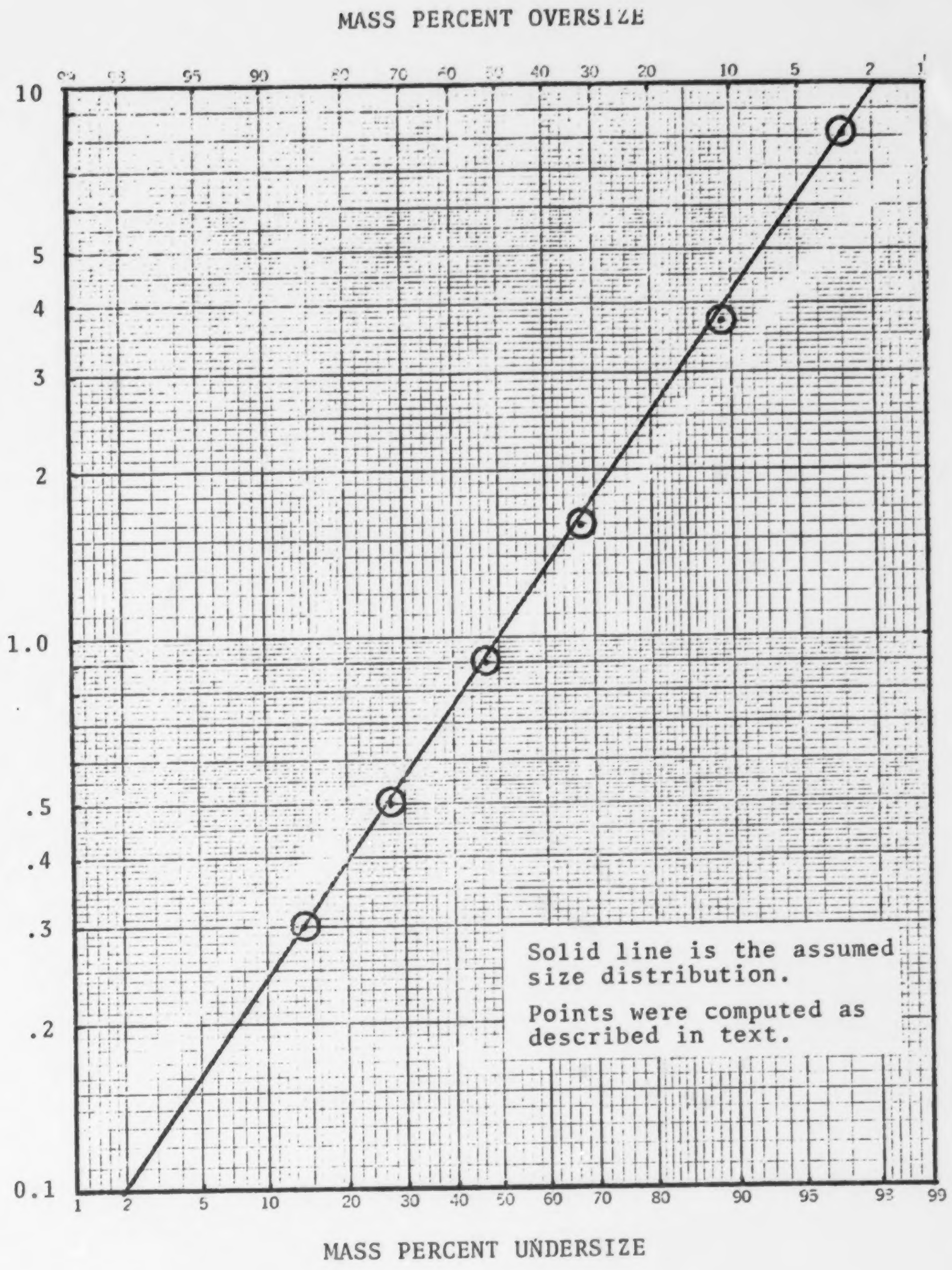
To illustrate the performance of a cascade impactor, assume that 100 mg of particles are carried into the impactor by a gas stream at 1 CFM flow rate. Also assume that the size distribution is log normal, that the mass median diameter (d_{pg}) is 1.0 μmA and that the geometric standard deviation (σ_g) is 3.0. The line in Figure 9.3 represents this size distribution on a log-probability graph.

The performance of each stage of the impactor can be predicted from Figure 9.1 and the results presented in Figure 9.4, a plot of mg of particles suspended in the gas versus particle aerodynamic diameter. The solid line represents the original size distribution (corresponding to Figure 9.3)and the dashed lines represent the particle suspension passing through, or penetrating, the various impactor stages. The first stage is neglected because its cut diameter is so high that a negligible amount of particles would collect there. Stage No. 2 has an 8.1 μmA cut point and it would collect 3% of the particles (i.e., 3 mg out of 100 mg). Stage No. 3 has a 3.8 μmA cut point and it would collect about 8.0 mg while allowing 89 mg to penetrate as the "undersize" fraction.

The mass of particles collected on a stage is very nearly the same as would be collected if the stage had a perfectly sharp size selection. For example, Stage No. 5 has a cut diameter of 0.9 μmA and since 46% of the particle mass is smaller than 0.9 μmA, a stage which collects everything larger than the cut size and nothing smaller would allow 46 mg to penetrate.

A computation of the performance of Stage No. 5 integrated over the example size distribution resulted in the prediction that about 46.3 mg would penetrate. It has been shown that the use of cut diameter to characterize the collection efficiency of an impaction stage will give results within a few percent of the correct answer so long as the geometric standard deviation is larger than about 2.0.

The stage cut diameters masses collected, and other data for this example are shown in Table 9.1, below. Particles which penetrate the last stage (No. 7) are separated from the gas by means of a filter with an efficiency of substantially 100%.



Solid line is the assumed size distribution.

Points were computed as described in text.

Figure 9.3 Log Probability Plot

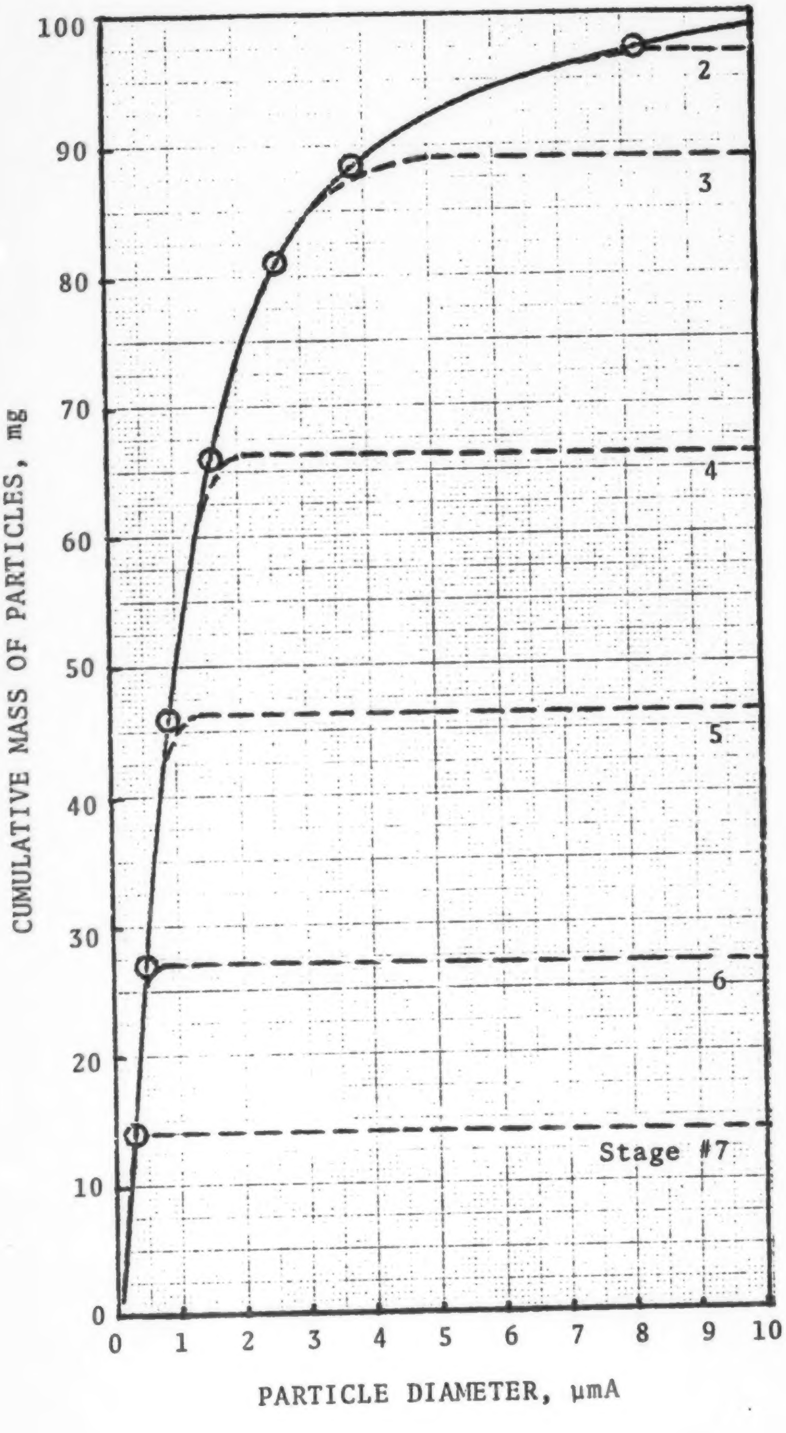


Figure 9.4 Particle Mass Penetrating Cascade Stages

TABLE 9.1

Stage No.	Mg on Stage	Stage Cut Dia. µmA	Cum. % Oversize	Cum. % Undersize
2	3.0	8.1	3.0	100.0
3	8.0	3.8	11.0	97.0
4	22.5	1.6	33.5	89.0
5	20.2	0.9	53.7	46.3
6	19.3	0.5	73.0	66.5
7	13.0	0.3	86.0	27.0
Filter	14.0	---	100.0	14.0
TOTAL	100.0			

Data Analysis

Once the stage "catches" have been measured, usually by weighing particle collection foils or papers, the data analysis is relatively simple. Generally the objective is to make a plot of particle diameter versus mass percent oversize or undersize and to represent the size distribution in terms of log-normal distribution parameters if possible. Thus it is necessary to do the following:

1. Add all of the stage and filter collection weights to get the total particle mass collected.
2. Compute either:
    a. Cumulative percent collected as the gas flows through succeeding stages. This is "percent oversize".
    b. Cumulative percent penetrating as the gas flows through succeeding stages. This is "percent undersize".
3. Compute the cut diameters for the impactor stages, taking into account gas viscosity (or temperature) and gas sampling flow rate.

At this point one can plot the data on log-probability paper or in any other format desired. To avoid confusion about what cut diameter is associated with what cumulative mass percentage, bear in mind the following points:

1. The cumulative mass collected by a stage and all those preceding it (i.e., through which the gas flowed before reaching this stage) represents the particles the same size and larger than the stage cut diameter.

2. The cumulative mass penetrating a stage and all those preceding it represents the particles smaller than the stage cut diameter.

To illustrate, consider the Stage No. 5 given in Table 9.1. The cumulative mass percent oversize is 53.7%, which represents the particle size class which is 0.9 μmA diameter or larger. One can plot this point on Figure 9.3, which has a percent oversize scale on top of the plot and a corresponding percent undersize scale on the bottom. The point is, as noted before, very close to the assumed size distribution that was the basis for this illustration.

Data points for the example are plotted on Figure 9.3. It is obvious that the process of fractionating the particles through a hypothetical cascade impactor and using the cut diameter method of data analysis has not introduced much error. In practice, the weighing errors and other experimental errors will overshadow any errors introduced by the cut diameter approximation to define stage efficiency.

EVALUATION OF A PARTICULATE SCRUBBER ON A FULL SCALE UTILITY BOILER

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EVALUATION OF A PARTICULATE SCRUBBER ON A FULL SCALE UTILITY BOILER

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Abstract

The paper presents results of a performance test and engineering analysis of a mobile-bed scrubber on a full-scale coal-fired utility boiler. The scrubber nominally operated at the design particulate removal efficiency of 95 percent, but the concentration of submicron particles was greatly influenced by mist entrainment. The entrainment resulted in a difference of aerosol penetration through the scrubber as a function of elemental composition and outlet submicron particle concentration independent of pressure drop through the scrubber. The engineering analysis showed that the 1972-installed cost was $29/kw and the annual operating cost is 0.5 mills/kwh (75 percent availability). An initial decline in scrubber availability after start-up resulted from now-corrected minor design problems. Steadily improving reliability is attributed to the utility's providing maintenance and solving operating problems.

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I. INTRODUCTION

Although scrubbers are commonly used for particulate control in many industries, they are not commonly used in the utility industry. Because scrubbers are not sensitive to dust resistivity, they may provide an acceptable means of controlling high-resistivity fly ash from combustion of low sulfur coal. In order to assess the potential of scrubbers for particulate control in the electric power industry, Industrial Environmental Research Laboratory-Research Triangle Park (IERL-RTP), sponsored an extensive evaluation of the TCA[*] scrubber at Public Service Company of Colorado's Cherokee Power Plant. A summary of the performance tests and engineering analysis of the evaluation is presented in this paper. Details of the tests and test results are reported by Ensor et al. (1975).

II. DESCRIPTION OF SYSTEM

Power Plant

The Cherokee Power Plant is located in north Denver, Colorado, near the South Platte River. The #3 unit has a nameplate rating of 150 mw. The gas cleaning equipment consists of mechanical collectors, an electrostatic precipitator, and a wet scrubber arranged in series.

The unit normally burns coal or natural gas. The coal is mined in western Colorado and has a sulfur content of about 0.5 percent, ash content of 9 percent, and a heating value of 2.492×10^7 J/kg (10,712 BTU per pound).

Scrubber

The Model 6700 Turbulent Contact Absorber Scrubber, designed by UOP, Air Correction Division, was installed in 1972. A diagram of the system is shown in Figure 1. The flue gas from the precipitator passes into two parallel induced draft fans. A bypass damper is used to direct the flue gas either into the stack or into the scrubber. The flue gas (under design conditions), 17,000 AM^3/min, at 137°C (610,000 ACFM at 280°F), enters the booster fans to offset the pressure drop through the scrubber. In the presaturator, 1440 ℓ/min (380 GPM) of makeup water is sprayed into the gas to reduce the temperature to approximately 52°C (125°F).

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[*] - Mention of manufacturers or brand names does not imply endorsement or recommendation for use by EPA.

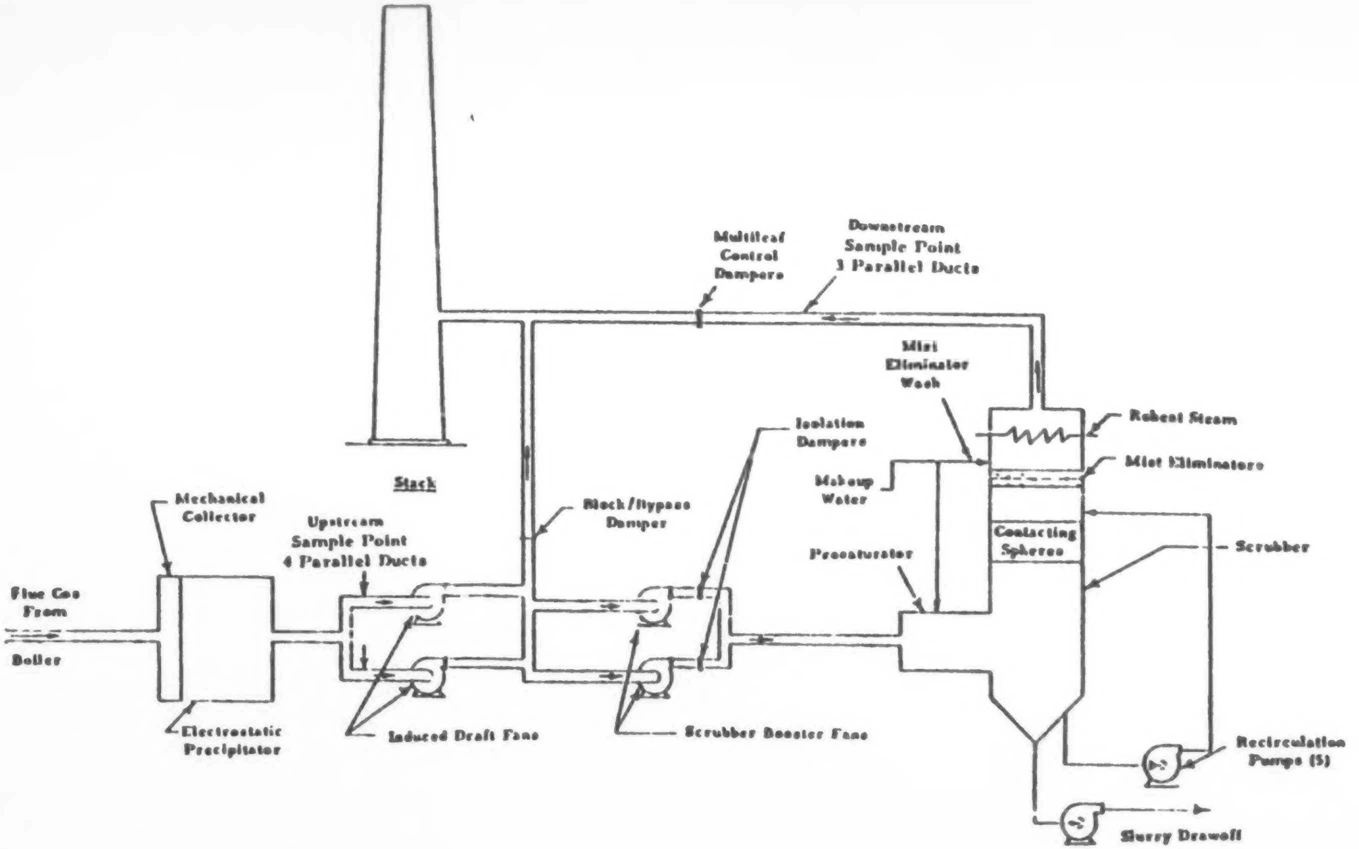

Fig. 1. SCRUBBER SYSTEM

From the presaturator, the gas enters the scrubber. The scrubber consists of three stages of fluidized beds packed with 3.8 cm (1.5 in.) diameter plastic balls arranged into three separated parallel scrubber sections. The two outer sections each handle 20 percent of the flow, while the center section handles the remaining 60 percent. All three sections can operate independently to provide flexibility of operation.

The scrubbed gas passes through chevron-type mist eliminators made of fiberglass reinforced plastic where entrained droplets are removed. The mist eliminators are sprayed once a work-shift from the top to prevent accumulation of solids.

The gas is then heated by steam coils to 85°C (185°F) before entering the stack to prevent corrosion of the stack and duct work and to provide plume buoyancy after discharge into the atmosphere. The steam coils are equipped with two sets of soot blowers to remove fly ash from the heat transfer surfaces.

The scrubber was designed to the following specifications (Raban, 1974):

Gas Flow	1,036,000 AM^3/hr, 610,000 ACFM 138°C (280°F)
Liquid-to-Gas Ratio	$7.4\ \ell/m^3$ (55 gal/1000 ft^3)
ΔP	30.5 cm H_2O (12 in. H_2O)
Inlet Particulate Concentration	0.92 gm/m^3 (0.40 grains/SCF)
Outlet Particulate Concentration	0.046 gm/m^3 (0.02 grains/SCF)
Efficiency of Particulate Removal	95 percent

III. MEASUREMENT TECHNIQUES

A total of three measurement techniques were used during the tests to determine particle collection efficiency. These were: (1) diffusional techniques using condensation nuclei counters and screen diffusion batteries for determining concentration and size distribution on a number basis for particles having diameters less than approximately 0.2 μm; (2) inertial techniques using Meteorology Research, Inc. (MRI), cascade impactors for determining concentrations and size distributions on a mass basis for particles having diameters between approximately 0.3 µm and 20 µm, and (3) standard EPA Method 5 mass train measurements for determining outlet mass loadings. Details of the various instruments and test procedures are reported by Ensor et al.(1975).

Two Meteorology Research, Inc. (MRI), Plant Process Visiometers (PPV) were installed--one at the scrubber inlet and the other at the outlet. These instruments were used as real time monitors to detect upsets and to determine in-stack opacity.

Impactor runs were conducted to obtain samples for elemental analysis by Ion-Excited X-Ray Analysis.

All impactor runs were conducted using Apiezon L high vacuum grease as a particle collection substrate. The Apiezon L was applied to light-weight collection discs. All weighings were conducted using a Cahn 400 analytical balance with a precision of 0.01 mg. Use of the Cahn balance and light weight substrates was essential to reduce errors due to weighing.

The test program was divided into three phases in order to systematically evaluate the scrubber. The test phases were:

- Phase 1 - Evaluation of the distribution of aerosol mass concentration at the inlet. With the limited resources in manpower and equipment, the evaluation of the source was attacked one aspect at a time. All four inlet ducts were sampled at the same time. The cascade impactor trains were supplemented with in-stack filter probes to establish the consistency of the concentration.

- Phase 2 - Evaluation of outlet distribution of aerosol mass concentration. Similar parallel tests were used on the outlet of the scrubber to indicate the variation in the sections.

- Phase 3 - Simultaneous inlet-outlet tests for scrubber efficiency. The final phase of the field program was the simultaneous inlet-outlet tests with both primary and secondary tests.

The tests disclosed several problems with the scrubber which are discussed by Ensor et al.(1975).

IV. SCRUBBER EFFICIENCY RESULTS

The Phase 3 tests were used to compute the efficiency of the scrubber. Sufficient data were taken to allow selection of the runs taken during steady-state plant conditions. The data used in the efficiency determinations are summarized in Table 1. The diffusion battery tests were restricted to section B at the outlet because of the relatively nominal behavior of that section of the scrubber.

Total Mass Collection Efficiency

The mass collection efficiency was computed using only the cascade impactor data. The results were summarized along with some control room data in Table 1. The pressure drop across the system was reported to indicate the magnitude of the effect of the existence of the air reheaters. The outlet gas flow was the total for the system as determined from the velocity traverses. The lack of correlation between efficiency and the pressure drop was an unexpected result.

The average efficiency of the December tests was 92 percent, slightly lower than the design efficiency of 95 percent.

Mass Penetration as a Function of Particle Diameter

The particle mass penetration as a function of particle diameter was determined for three days. These data are shown in Figures 2 to 4.

Generally, the size distributions and penetrations were consistent, though loadings fluctuated considerably. The maximum penetration (minimum efficiency) was achieved in the region of 0.2 μm actual diameter, similar to results for another scrubber as reported by Sparks et al.(1974).

Elemental Chemical Analysis

The measurement of the concentrations of a number of elements was intended mainly to provide data to aid in the interpretation of the scrubber performance. It was not planned to perform elemental balances or determine emission factors. The samples were taken during December 10-11, 1974. The overall penetration of various elements is shown in Table 2. The penetration as a function of particle diameter for various elements is shown in Figure 5.

Table 1. SCRUBBER COLLECTION EFFICIENCY

				SECTION A				SECTION B				SECTION C			
DATE 1974	LOAD mw	O_2 Percent	OUTLET GAS FLOW ACTUAL m^3/hr	ΔP SYSTEM cmH_2O	ΔP BED cmH_2O	ΔP MIST ELIMIN. cmH_2O	EFF. %	ΔP SYSTEM cmH_2O	ΔP BED cmH_2O	ΔP MIST ELIMIN. cmH_2O	EFF. %	ΔP SYSTEM cmH_2O	ΔP BED cmH_2O	ΔP MIST ELIMIN. cmH_2O	EFF. %
11/20	165	3.6	[a]	41	9.9	0.76	ND	45	25	2.5	84.7	46	24	8.3	ND
11/21	164	3.4	[a]	39	9.6	0.76	ND	43	18	1.8	89.9	44	20	5.1	ND
12/10[b]	157	3.4	9.47×10^6	36	15.2	1.7	96.3	41	20.8	2.5	92.6	41	18.5	3.8	86.9
12/11	160	3.0	10.2×10^6	38	14.7	1.5	96.4	42	22.1	3.2	93.2	44	22.4	2.5	96.7
12/12	160	2.6	8.78×10^6	38	14.7	1.8	79.6	44	22.9	2.5	93.1	46	24.1	3.8	92.1

[a] Full velocity traverses were not taken.

[b] The control room data were incomplete. Interviews, data from other days and the log book were used to supplement available information.

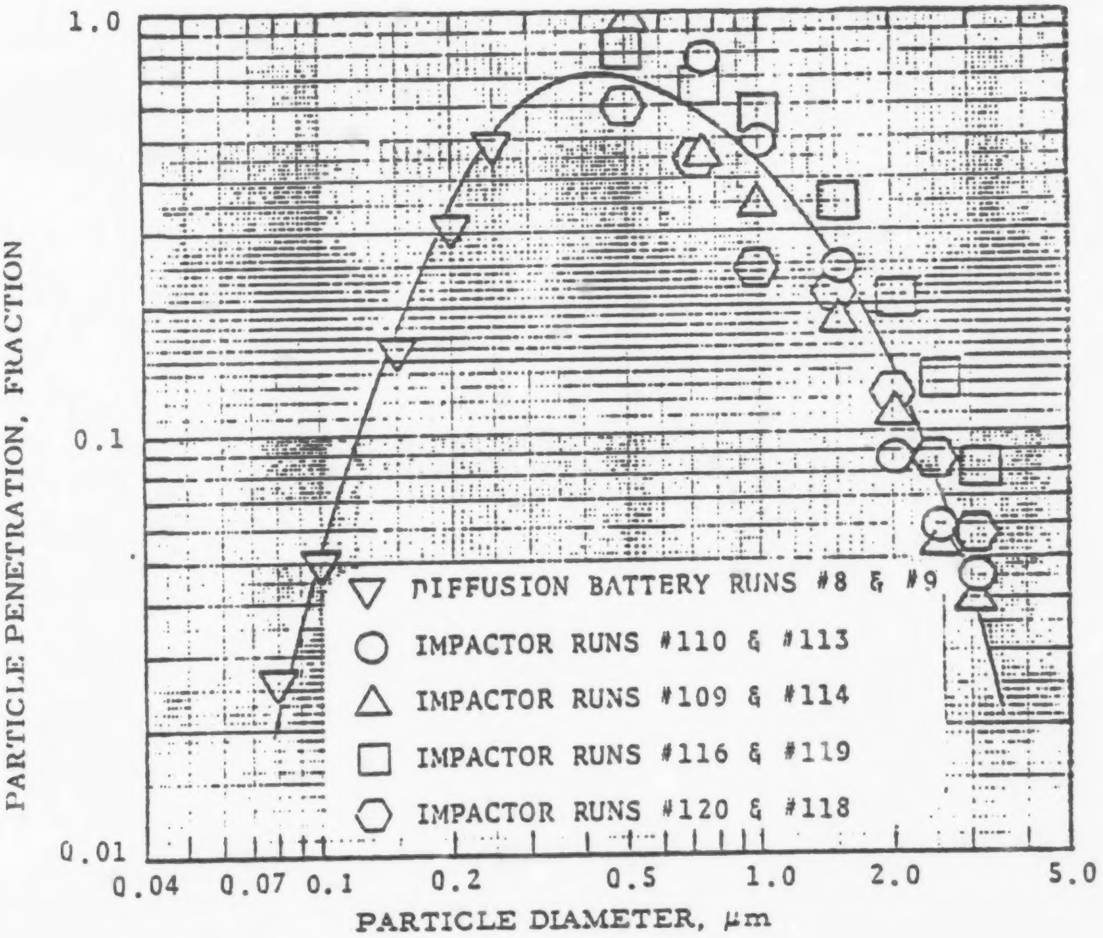

Fig. 2. COMBINED PENETRATIONS FOR DIFFUSION BATTERY AND CASCADE IMPACTOR (December 10, 1974)

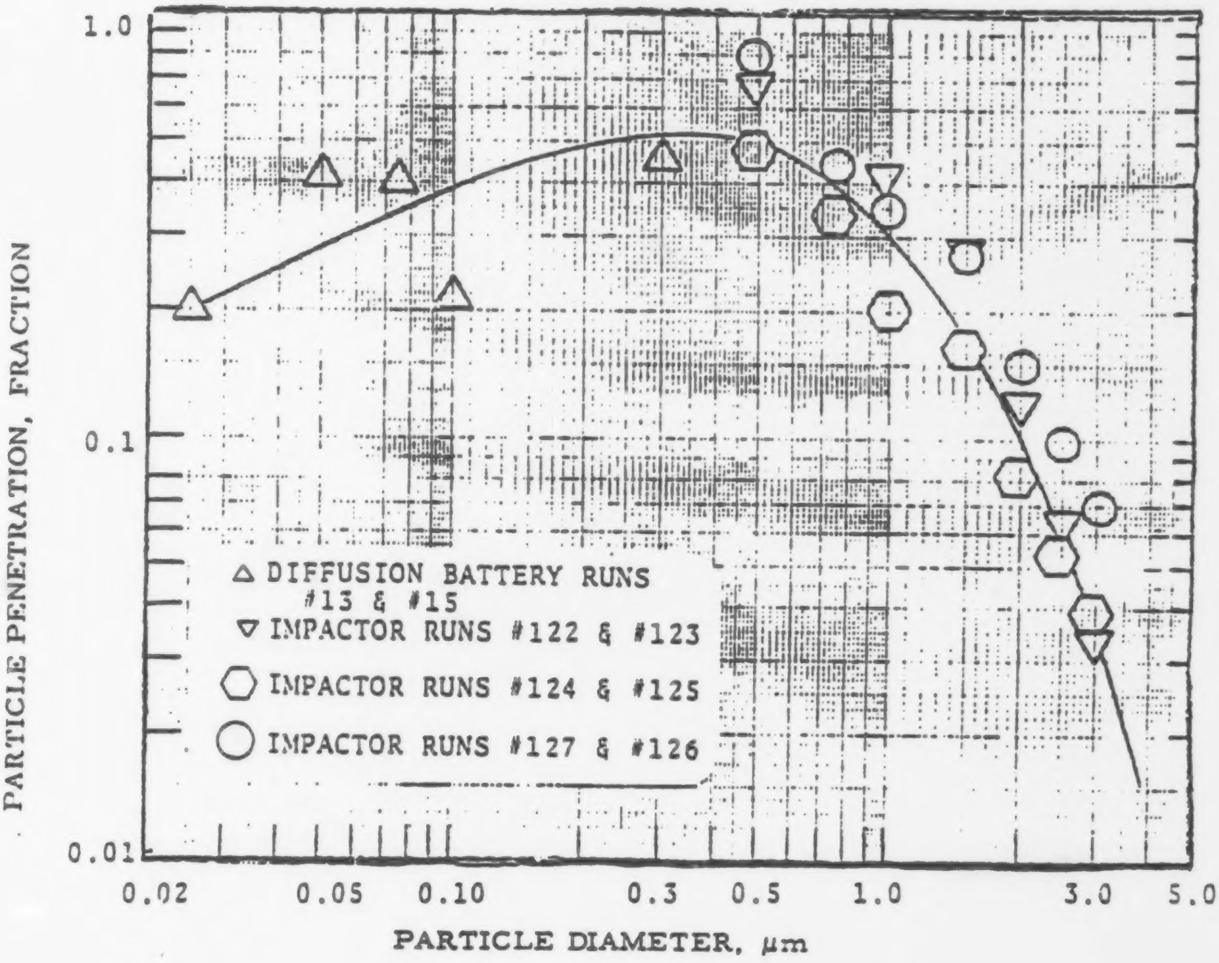

Fig. 3. COMBINED PENETRATIONS OF DIFFUSION BATTERY AND CASCADE IMPACTOR (December 11, 1974)

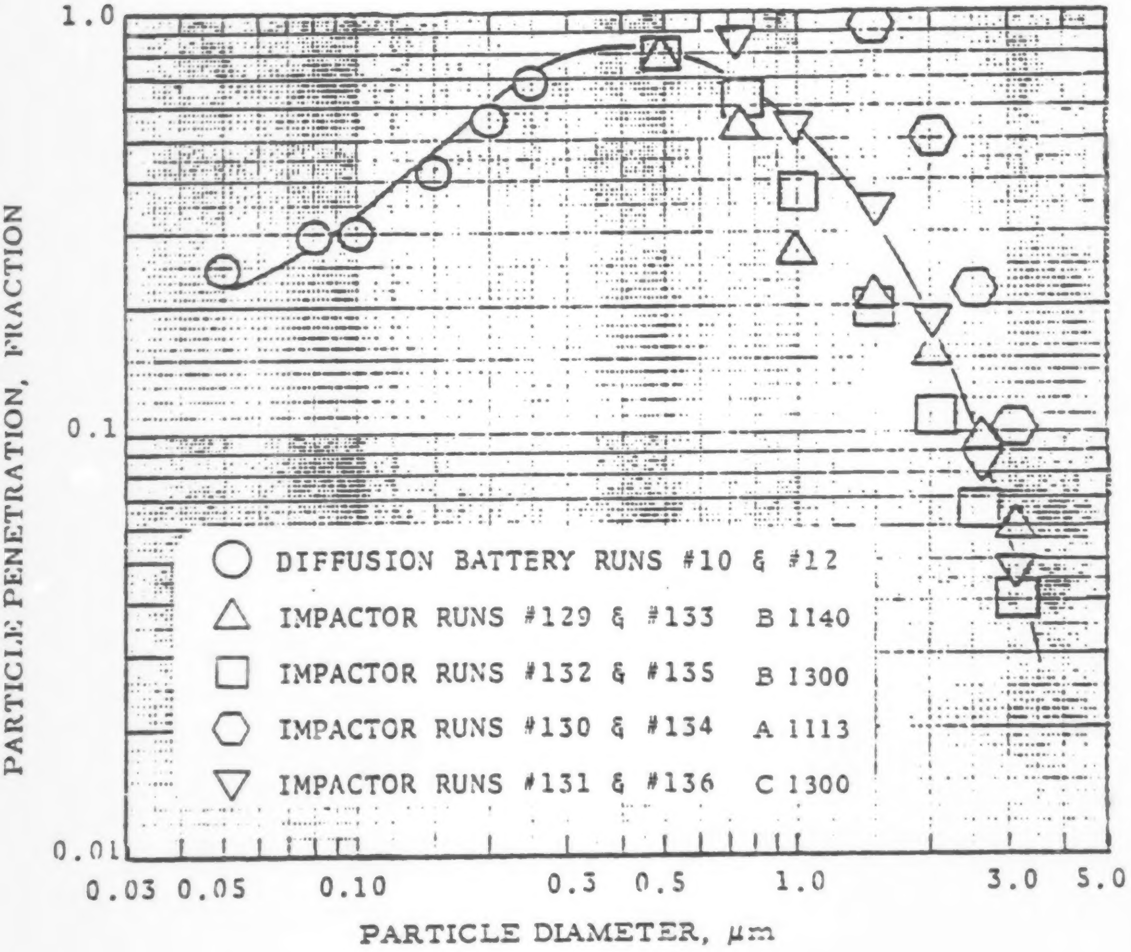

Fig. 4. COMBINED PENETRATIONS FOR DIFFUSION BATTERY AND CASCADE IMPACTOR (December 12, 1974)

Table 2. PENETRATION OF THE ELEMENTS THROUGH THE SCRUBBER FOR DECEMBER 10, 1974

Element	Penetration	Average Outlet Concentrations Micrograms/DSm^3 [b]
Al	0.029	326
Si	0.033	658
S	3.4 [a]	1030
K	0.043	50
Ca	0.059	508
Ti	0.073	96
V	0.14	27
Cr	1.10 [a]	57
Fe	0.18	1500
Ni	0.95	33
Cu	2.9 [a]	668
Zn	1.5 [a]	501
Br	0.28	5.7
Pb	0.64	120
For All Elements	0.108	
Total Mass	0.074	

[a] Penetrations greater than 1 indicate generation of particles in flue gas by evaporation of the scrubber liquor.

[b] Dry Standard, 21.1°C, 760 mm Hg.

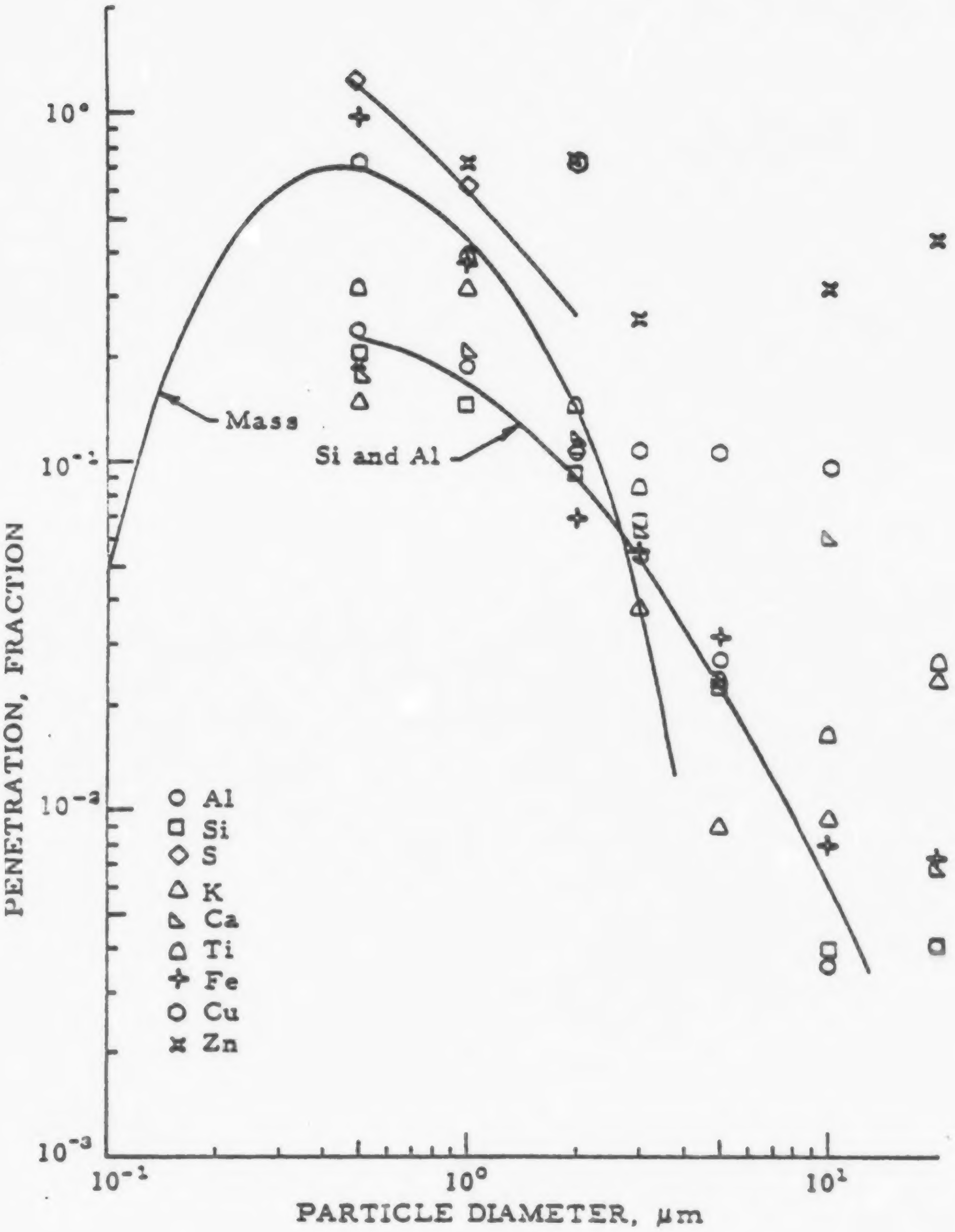

Fig. 5. SCRUBBER PENETRATIONS FOR SELECTED ELEMENTS

V. DISCUSSION

The scrubber performance results were analyzed in depth for trends and to allow comparison to data in the literature.

Examination of the particle penetration data for each simultaneous pair of inlet and outlet runs shows a wide range of results, as summarized in the following table.

TABLE 3. PARTICLE PENETRATION SUMMARY

RUN NO.		SECTION	d_{pc},[a] μm	ΔP cm H_2O	F. out / F. in[b]
IN	OUT				
56	53	B	1.5	23.0	0.31
55	57	B	1.1	23.0	0.81
60	61	B	0.8	23.0	1.2
114	109	B	0.8	23.0	2.9
119	116	B	1.0	23.0	0.39
121	123	B	0.8	23.0	0.55
125	127	B	< 0.5	23.0	1.2
133	129	B	0.8	23.0	2.0
135	132	B	0.8	23.0	0.47
124	125	B	0.6	23.0	3.1
113	110	A	1.0	--	0.81
127	126	A	0.8	15.0	0.41
134	130	A	2.0	9.0	6.7
115	119	C	< 0.5	25.0	0.76
120	118	C	0.6	--	1.0
123	122	C	0.7	25.0	1.1
136	131	C	1.1	24.0	3.5

[a] Scrubber cut (50 percent penetration) diameter.

[b] Ratio of outlet filter to inlet filter particle concentrations (mg/DS m^3)/(mg/DS m^3).

The scrubber performance cut diameter (i.e., particle diameter at 50 percent penetration), used to characterize efficiency, varied from less than 0.6 µm to 2.0 µm. No pattern of correlation between cut diameter and other parameters such as pressure drop was found. Thus, the variation was due to a combination of system fluctuations and measurement errors.

Penetrations found in this study are much higher than those reported in a previous study of the same scrubber (Calvert et al. 1974) and of another mobile bed scrubber on a coal-fired power plant (Statnick and Drehmel, 1974). For example, Calvert et al. reported a cut diameter of about 0.35 µm and a penetration at 1.0 µm of about 0.09. Statnick and Drehmel reported that, for 25 cm of water scrubber pressure drop, penetration did not exceed 0.05 for any particle size and was about 0.02 for 1.0 µm particle diameter. These and some other points are compared in the following table with representative results from the present study.

TABLE 4. COMPARISON OF MOBILE BED STUDIES

INVESTIGATORS	d_{pc}, µm	PENETRATION AT PARTICLE DIAMETER SHOWN 0.5 µm	0.8 µm	1.0 µm	2.0 µm
Present	0.8	0.8	0.5	0.4	0.15
Calvert et al (1974)	0.35	0.3	0.15	0.08	0.02
Statnick and Drehmel (1974)	----	0.07	0.03	0.02	0.002

Approximate mean values for all runs, exclusive of 56/53 and 134/130.

There is evidence that the high outlet particle concentration measured is due to entrainment from the scrubber. Facts which indicate the presence of outlet particles introduced by entrainment are as follows:

1. Outlet filter loadings are higher than inlets, based on the same gas sample volume, for many of the tests.

2. Penetrations in the present study are higher than those found by others.

3. Variations in penetration are not related to gas flow rate, pressure drop, or other known parameters, but can be attributed to variable entrainment.

4. Penetrations for scrubber section "A" are the same as for section "B" despite the gas pressure drop for "A" being about half that for "B".

5. Reheater and entrainment separator operating problems occurred during the test.

6. Elemental analyses of inlet and outlet particles showed an apparent "generation" of particles containing soluble elements as described in the previous section.

The overall conclusion is that the scrubber performance data obtained in this study are specific for the operating factors and scrubber condition which existed during the test period. It is not possible to establish a general mobile bed scrubber performance model from these data because of the overshadowing and undefined influence of liquid entrainment. It appears that both scrubber performance and reliability could be improved by improved entrainment separation.

VI. ENGINEERING ANALYSIS

The objective of the engineering analysis was to assemble the following information:

- Capital costs
- Operating costs
- Major maintenance problems
- Scrubber reliability
- Estimated cost required to minimize operating problems.

Capital Costs

The total installed cost in 1972 for the scrubber was $4,400,000. Based on the boiler nameplate rating of 150 mw, the cost is $29/kw or, based on the rated gas volume (see Section I), is $4.18/1000 m^3/hr ($7.10/1000 ACFM). The detailed cost itemization is given by Ensor et al. (1975). In 1975 dollars, the scrubber would cost $5,800,000.

Operating Costs

The total operating costs are approximately $495,000/year (fourth quarter 1973 and the first three quarters of 1974) based on 75 percent availability of the scrubber or 0.50 mills/kwh.

Maintenance Problems

The scrubber had a number of maintenance problems, many of which were solved during start-up and operation. The problems mentioned here are the more persistent ones which have defied solution. The maintenance problems are discussed in detail by Ensor et al. (1975). These include:

1. Breakage of Mobile Bed Contactors - The plastic mobile bed contactors have been a chronic problem from breakage from wear. The desired lifetime of the spheres is 8000 hours, however, normally a lifetime of only 6000 hours was experienced.

   However, damage to the pump liners and plugging of the nozzles can result from the fragments entering the liquor recycle piping. Screens in the scrubber hopper have eliminated this problem with the penalty of adding screen cleaning to the required maintenance. Public Service Company of Colorado has tested a number of different packings and is currently using polyethylene spheres in the scrubber.

2. Migration of Mobile Bed Contactors - The mobile bed contractors will also migrate from one section of the scrubber to another if an opening the width of a contactor exists in the partitions. Poor distribution of the contactors causes channeling of flue gas and a reduction in particle collection efficiency.

3. Guillotine Dampers - The guillotine isolation dampers have caused problems due to breakage of the damper when closing against a buildup of fly ash and leakage.

4. Recirculation Pumps - The recirculation pumps have been a source of problems in the past due to mechanical failure. A new pump has been tested by Public Service Company of Colorado and has been providing good service.

5. Reheater Section - The scrubbed gases are heated by direct contact with three banks of steam coils. These coils are susceptible to pluggage and corrosion in the wet flue gas scrubber discharge. Addition of a second set of soot blowers and drying ash during periods of the scrubber shutdown with heavy soot blowing has had minimal effect. Corrosion problems have rendered the heaters inoperable. Both upper and lower layer reheat coils have been removed.

6. Weather-Related Problems - The freezing of lines has always been a problem during cold weather. All lines must be heat-traced and drained when the scrubber is shut down.

   Another solution is to enclose the scrubber in a weatherproof building. The major problem with this is the possibility of leakage of flue gas into the structure.

Scrubber Reliability

For this study, availability of the scrubber was defined as:

$$\text{Availability} = \frac{\text{hours of scrubber operation}^{*} - \text{hours boiler was burning 100\% gas}}{\text{hours of boiler operation} - \text{hours boiler was burning 100\% gas}}$$

The reason that the hours the boiler was burning 100 percent gas was subtracted from the total hours of scrubber and boiler operation is that the scrubber is normally either left running with only water systems operating or completely shut down during periods of 100 percent gas burning operation

It is interesting to note that the scrubber was out of service a majority of the time during August and September 1974, when the unit returned to mixed fuel burning. The reason for this is that after long periods of idleness, due to either maintenance or 100 percent gas burning, the scrubber

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[*] - Hours of scrubber operation is related directly to the time the boiler was operating.

usually experienced minor start-up problems. When start-up troubles with the scrubber were encountered, every effort was made to use any available gas on the unit. Consequently the scrubber had a poor percentage availability for the months involved, but the overall availability was not affected very much.

Values of availability are given for two capacity levels: 100 percent of capacity and 80 percent of capacity or greater. There is a significant amount of time that the scrubber operated at 80 percent of capacity. For example, in January of 1974, although the scrubber only operated at 100 percent of capacity for 175 hours, it did operate at 80 percent of capacity for 547 hours. Although the scrubber is not treating all of the boiler flue gas, it is felt that some credit should be given the scrubber for handling most of the flue gas. In most cases, running at 80 percent of capacity will probably allow PSCC to meet particulate emission standards.

The accumulative percent availability to date reflects the total hours of scrubber operation since start-up. This value is plotted by month in Figure 6. This figure shows that after an initial drop, the availability of the scrubber has been increasing steadily. This is typical of other PSCC scrubber operations. The initial drop is attributed to the many minor design problems which arose soon after start-up. The following increase in availability is attributed to the solving of these problems and the increased operating knowledge which was gained through operating experience.

As of November 30, 1974, the scrubber had operated at 100 percent capacity 59.9 percent of the time and at 80 percent capacity or greater 70.9 percent of the time.

In order to more clearly see the increase in availability with time, after the first year when each new month was added to the log, only the previous 12 months availability was calculated. This value is plotted in Figure 7. Looking at the last month tabulated (November 1974), reveals that the scrubber operated at 100 percent of capacity 78.6 percent of the time and at 80 percent of capacity or greater 96.7 percent of the time during the previous year. The figure also shows that the scrubber had been operating at 80 percent of capacity or greater with over 90 percent availability since May of 1973.

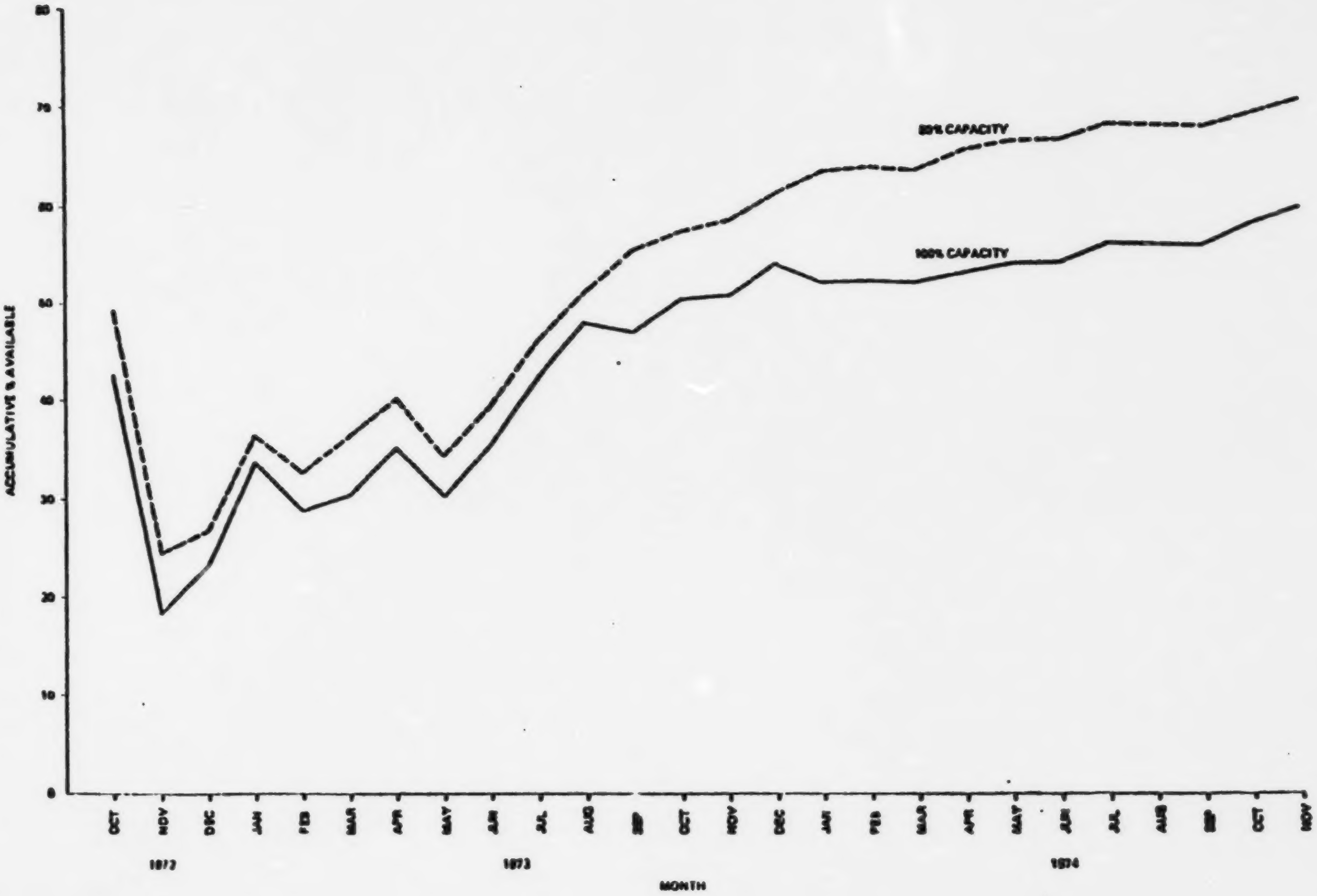

Fig. 6. CHEROKEE #3 SCRUBBER ACCUMULATIVE AVAILABILITY FROM START-UP TO NOVEMBER 1974.

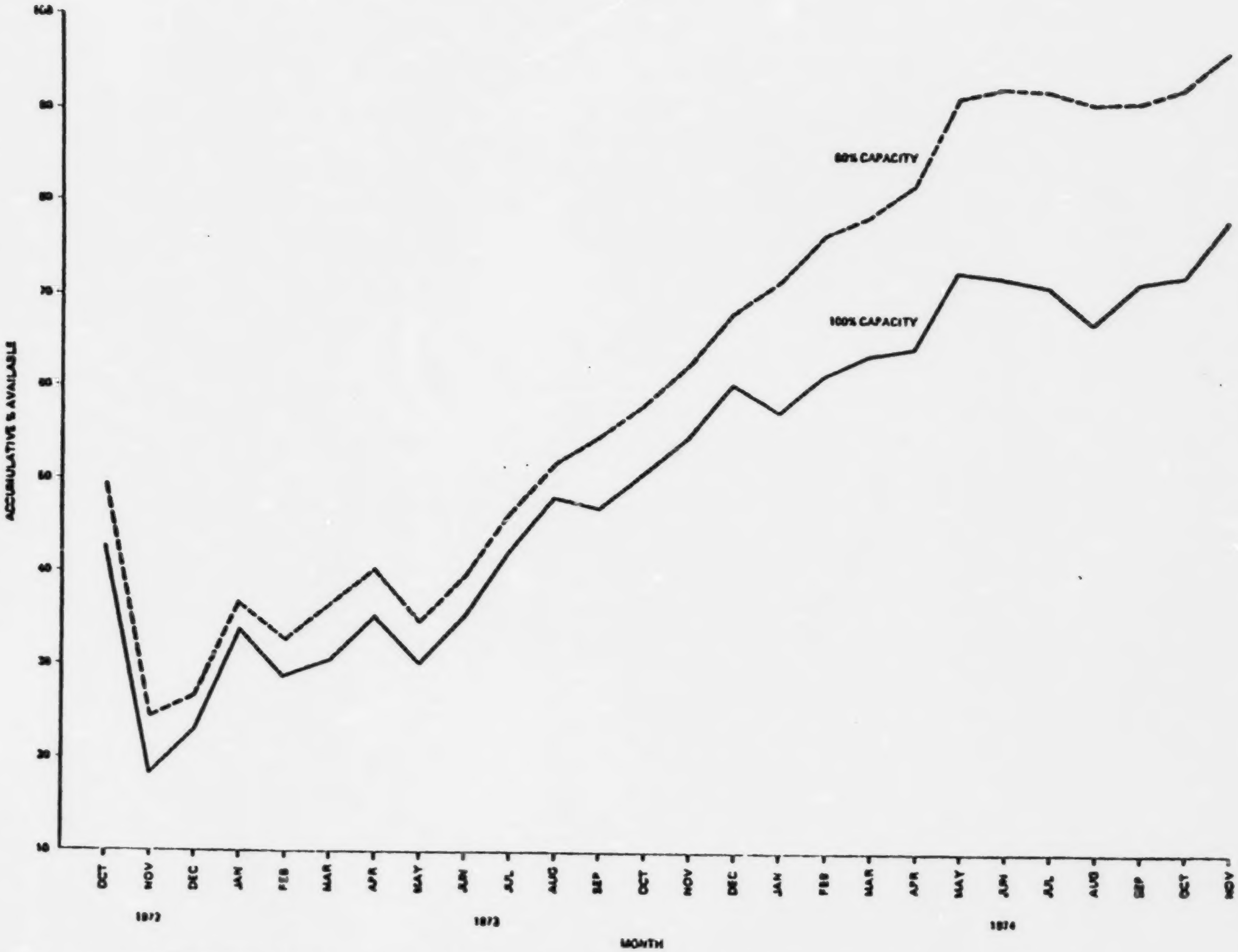

Fig. 7. CHEROKEE #3 SCRUBBER ACCUMULATIVE AVAILABILITY FROM START-UP TO NOVEMBER 1974 AVERAGED OVER THE PREVIOUS YEAR ON A RUNNING BASIS.

The feasibility of using the design and operating experience to improve availability of the scrubber was investigated. The goal was to increase the availability of the scrubber, realizing that the maintenance required may slightly increase due to the extra equipment. The suggested modifications are summarized below:

		Estimated Availability (Percent)
Identical Scrubber	\$5,800,000*	60-70
Extra 33-1/3 Percent Capacity Section	1,200,000	10-15
Indirect Reheat, Incremental	200,000	
Scrubber Enclosure	90,000	5-10
Miscellaneous Charges	80,000	
	\$7,370,000	75-95

The capital cost of \$7,370,000 is \$49/kw for a 150 mw unit.

The important aspect of maintaining and improving the availability of the scrubber system is the attitude of the plant operating and maintenance people in keeping the unit on line.

The steadily increasing availability of the Cherokee scrubber is an indication that the Public Service of Colorado is committed to solving the operating and maintenance problems.

VII. CONCLUSIONS

The performance data and engineering analysis indicate that scrubbers are capable of collecting fly ash from combustion of low sulfur coals. Capital and operating costs for a scrubber are likely to be higher than similar costs for a moderately sized electrostatic precipitator. However, in applications where the dust resisitivity requires a large-sized electrostatic precipitator, a scrubber system is likely to be cost competitive.

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* - 1975 dollars.

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COLLECTION MECHANISMS AT HIGH TEMPERATURE AND PRESSURE

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COLLECTION MECHANISMS AT HIGH TEMPERATURE AND PRESSURE

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ABSTRACT

The paper reports the results of a theoretical study of particle collection mechanisms at high temperature (1,100°C) and high pressure (15 atm). The effects of high temperature and pressure are presented for various collection mechanisms, including: inertial impaction, Brownian diffusion, electrical migration, thermophoresis, diffusiophoresis, centrifugal forces, gravitational forces, and particle agglomeration. Some examples of the effects of temperature and pressure on the performance of particulate control devices are also presented.

In general, high pressure reduces the effectiveness of particle collection mechanisms. For a given pressure, high temperature further reduces the effectiveness of collecting large particles, but improves the collection of particles smaller than about 0.1 to 0.5 μm in diameter. The beneficial effects of temperature are much more pronounced at lower pressures (near atmospheric).

COLLECTION MECHANISMS AT HIGH TEMPERATURE AND PRESSURE

INTRODUCTION

The removal of particles from high temperature and pressure gas streams is a problem which is encountered in many existing and proposed energy processes. One important example is the removal of particles from low-BTU coal gas before it is combusted and used to drive a combined-cycle gas turbine/steam turbine power generation system. The particles must be removed to protect the turbine blades as well as to satsify the particulate emissions standards, and must be removed with a minimum loss of pressure and temperature. High temperature particulate removal is also of interest in other processes, such as some metallurgical processes, where heat recovery is not economical and expensive cooling equipment is required to reduce the effluent gas temperature to a temperature suitable for conventional particulate control. The effects of high temperature and pressure on particle collection have been studied by many authors including Thring and Strauss[1], Strauss and Lancaster[2], Calvert, et al.[3], and Rao, et al.[4]. The present study represents a review and extension of the material presented by previous authors.

All types of devices used for particle collection can be characterized by a deposition velocity, "v_d", which can be related to the particle collection efficiency, "E", through the penetration, "Pt", as follows:

$$Pt = 1 - E = \exp\left(- \frac{v_d\ A_d}{Q_G}\right) \tag{1}$$

where

A_d = deposition area, cm^2
Q_G = volumetric flow rate, cm^3/s

The particle penetration is defined as:

$$Pt = \frac{\text{Outlet concentration}}{\text{Inlet concentration}} \qquad (2)$$

The deposition velocity for any collection mechanism depends on the force balance between the driving force (deposition force) and the resistance force of the gas. As will be shown later, the fluid resistance force is generally affected by high temperature and pressure much more than the driving force.

Table 1 lists the most important particle collection mechanisms for high temperature and pressure collection. The effects of temperature and pressure on these mechanisms are presented below. Some examples of the effects of temperature and pressure on the performance of specific devices are also presented.

Table 1. IMPORTANT COLLECTION MECHANISMS AT HIGH TEMPERATURE AND PRESSURE

1. Inertial impaction
2. Brownian diffusion
3. Electrical migration
4. Gravitational settling
5. Centrifugal separation
6. Thermophoresis
7. Diffusiophoresis
8. Particle size alteration
   a. thermal coagulation
   b. turbulent coagulation
   c. sonic agglomeration

FLUID RESISTANCE FORCE

The major difference between the collection of particles at normal conditions and at high temperature and pressure is in the fluid resistance force. The fluid resistance force is generally approximated by Stokes' law modified to allow for non-continuum slip flow effects.

$$F_r = \frac{3\pi \mu_G d_p u_o}{C'} \tag{3}$$

where

F_r = fluid resistance force, dynes
μ_G = fluid dynamic viscosity, g/cm-s
d_p = particle diameter, cm
u_o = relative velocity between the particle and the gas, cm/s
C' = Cunningham slip correction factor, dimensionless

The temperature and pressure dependence of equation (1) is contained in the terms "μ_G" and "C'". The viscosity of a gas increases with increasing temperature and at extreme pressures, viscosity increases slightly with pressure. This effect is not significant at pressures below about 20 atm.

The Cunningham slip correction factor may be calculated as:

$$C' = 1 + \frac{2\lambda}{d_p}\left[1.257 + 0.40 \exp\left(-1.1\, d_p/2\lambda\right)\right] \tag{4}$$

where

λ = mean free path of gas molecules, cm

The Cunningham slip correction factor is a function of temperature, pressure, and particle diameter, and is only important for small particles, high temperatures, and low pressures.

The particle deposition velocity for most collection mechanisms of interest is inversely proportional to the fluid resistance force, and therefore proportional to the ratio "C'/μ_G". The effects of high temperature and pressure on this ratio, plotted as a function of particle diameter, are illustrated in Figure 1. At atmospheric pressure, the ratio decreases with increasing temperature for particles larger than about 0.6 µm. At 15 atm pressure, the ratio decreases with temperature for all particle sizes considered. Therefore, the particle deposition velocity will generally be smaller at high temperature and pressure than at normal conditions.

COLLECTION MECHANISMS

Inertial Impaction

One of the most important mechanisms for collecting particles greater than a few tenths of a micrometer in diameter is inertial impaction. Inertial impaction is the collection of moving particles by impinging them on some target. The relative effect of inertial impaction for different particles and flow conditions may be characterized by the inertial impaction parameter, "K_p", defined as:

$$K_p = \frac{C' \rho_p d_p^2 u_o}{9 \mu_G d_c} \tag{5}$$

where

ρ_p = particle density, g/cm³
d_c = characteristic length for collector, cm

The inertial impaction parameter is equivalent to the ratio of the particle stopping distance, "x_s", to "d_c". The particle stopping distance is that distance the particle would travel before coming to rest if injected into a still gas at a velocity "u_o", when only the fluid resistance force acts on the particle. By considering the particle stopping distance divided by "u_o", the particle's inertia can be characterized by a relaxation time, "τ", defined as:

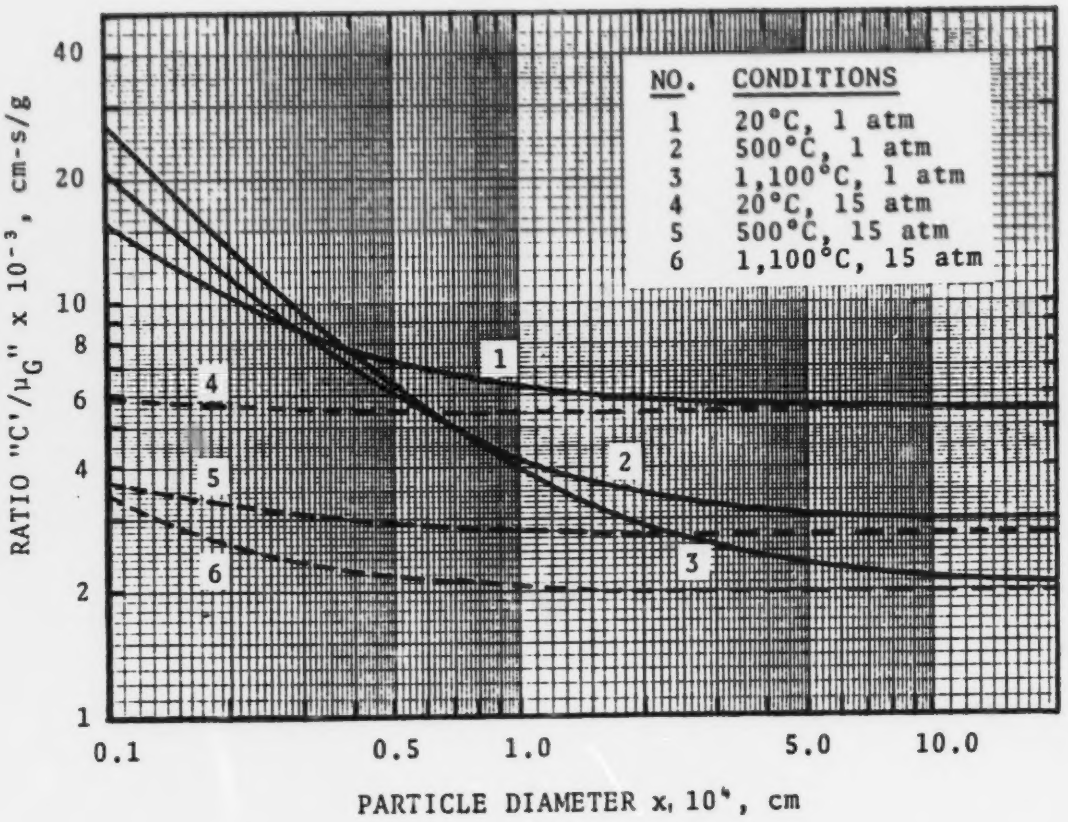

Figure 1. The effect of temperature and pressure on the ratio of the Cunningham slip correction factor to the dynamic viscosity of air.

$$\tau = \frac{x_s}{u_o} = \frac{K_p\ d_c}{u_o} = \frac{C'\ \rho_p\ d_p{}^2}{9\ \mu_G} \tag{6}$$

Figure 2 shows the effects of high temperature and pressure on the particle relaxation time plotted as a function of the compensated particle diameter (particle diameter times the square root of the particle density). For large particles, inertial impaction decreases with increasing temperature and pressure.

For small particles, high temperature and atmospheric pressure curve 2 indicates that inertial impaction begins to improve at very high temperature. High pressure effectively nullifies any beneficial effect of high temperature (curve 3).

Brownian Diffusion

Small particles can undergo significant Brownian motion resulting from the random bombardment of the particle by gas molecules. The rate of diffusion is characterized by the particle diffusivity, defined as:

$$D = \frac{C'\ k\ T}{3\pi\ \mu_G\ d_p} \tag{7}$$

where

D = particle diffusivity, cm²/s

k = Boltzman's constant, erg/°K

T = absolute temperature, °K

Figure 3 shows the effects of temperature and pressure on particle diffusivity as a function of compensated particle diameter. Smaller particles undergo higher rates of diffusion. High temperature increases the diffusivity for all particle sizes, however, this effect is greater for smaller particles. High pressure decreases the beneficial effect of temperature. The combined effect of high pressure and high temperature is a net increase in diffusivity.

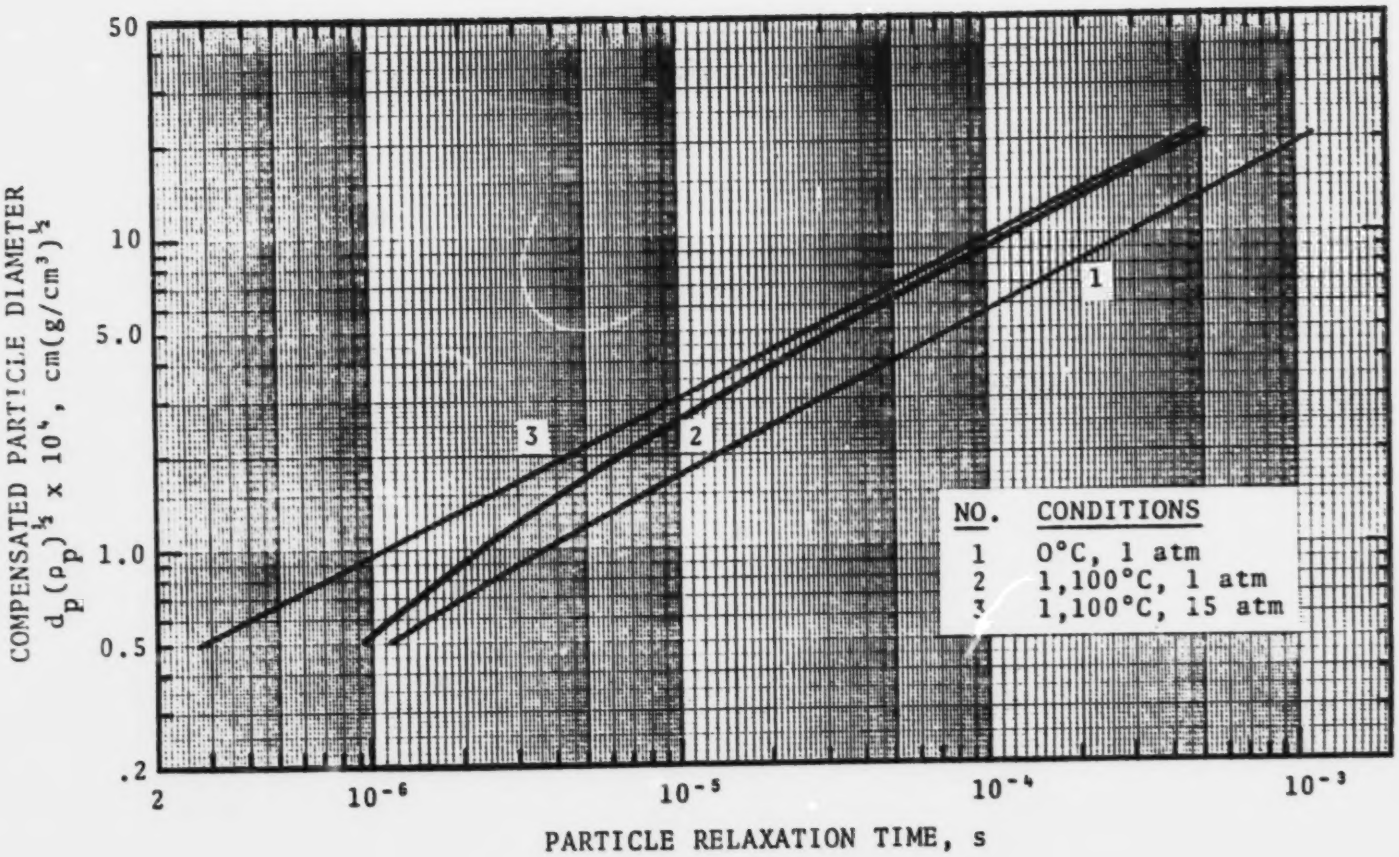

Figure 2. The effect of high temperature and pressure on particle inertia.

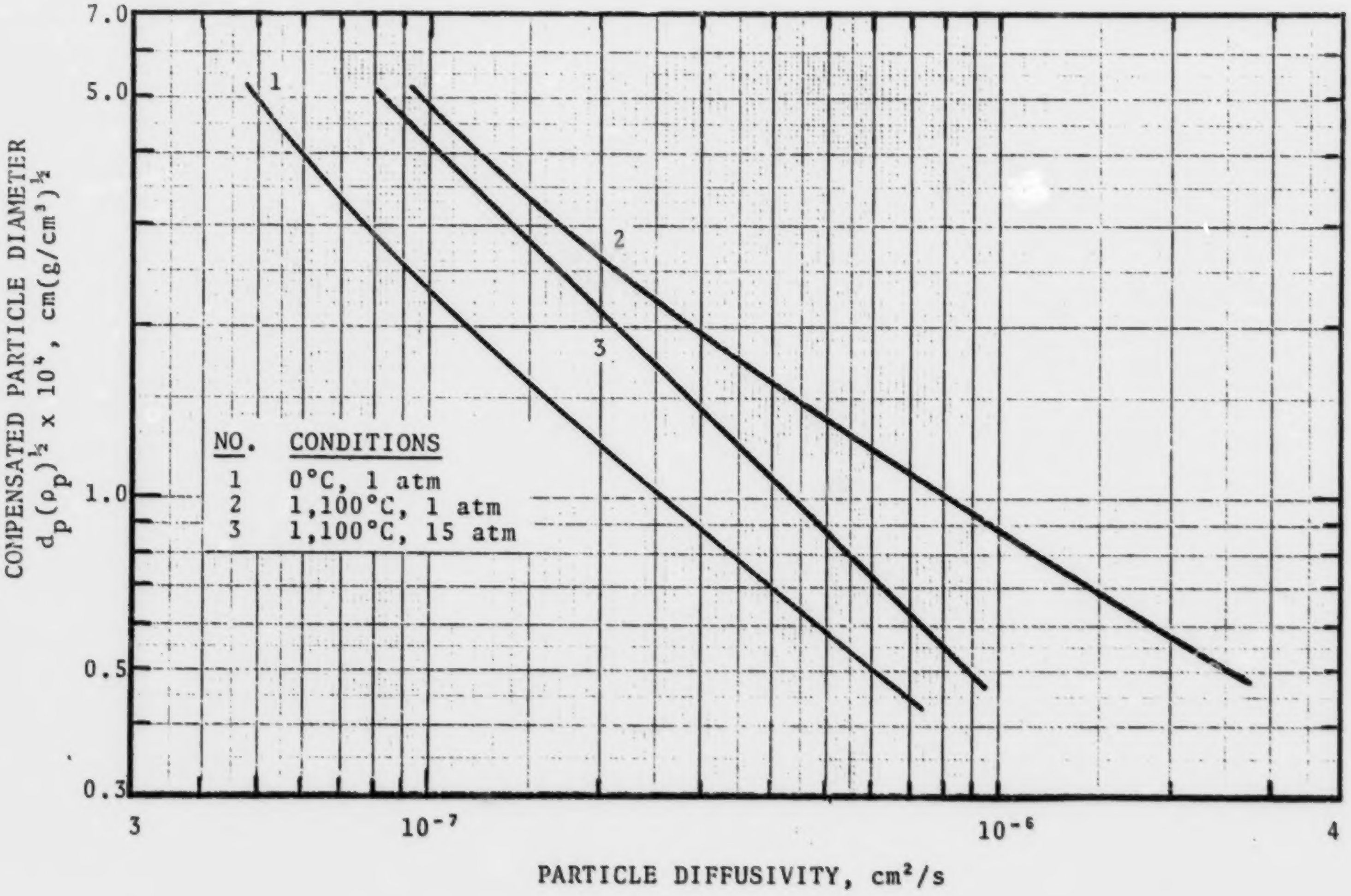

Figure 3. Particle diffusivity at high temperature and pressure.

The diffusional deposition velocity can be shown to be proportional to the square root of the diffusivity. That is,

$$u_{BD} \sim (D)^{1/2} \tag{8}$$

Using equation (1), the penetration can be related to diffusivity as:

$$\frac{\ell n\ Pt_{T,P}}{\ell n\ Pt_o} = \left(\frac{D_{T,P}}{D_o}\right)^{1/2} \tag{9}$$

where

$Pt_{T,P}$ = penetration at temperature, T, and Pressure, P
Pt_o = penetration at standard conditions
$D_{T,P}$ = diffusivity at "T" and "P", cm^2/s
D_o = diffusivity at standard conditions, cm^2/s

Figure 4 is a plot of the diffusivity ratio, "$D_{T,P}/D_o$", as a function of penetration for three assumed penetrations at standard conditions. For a given temperature and pressure, the diffusivity ratio can be obtained from Figure 3 and the effect of this diffusivity ratio on the penetration can be obtained from Figure 4.

Electrical Migration

The migration of a particle in an electric potential field is proportional to the field strength, the particle charge, and the fluid resistance force. Electrical migration is generally characterized by a deposition velocity (migration velocity) defined as:

$$u_e = \frac{C'\ q_p\ E}{3\pi\ \mu_G\ d_p}\ (10^7) \tag{10}$$

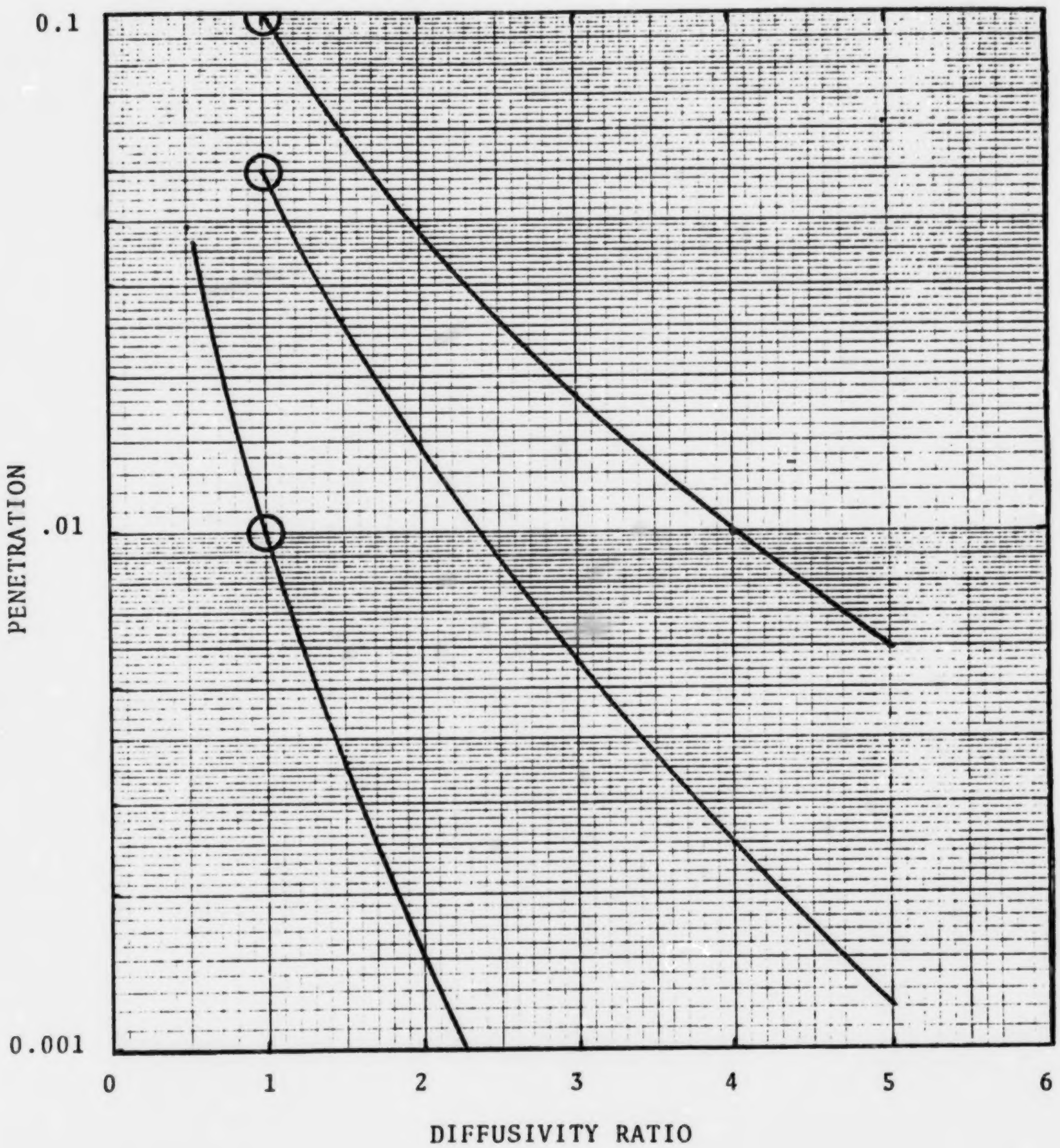

Figure 4. Penetration as a function of particle diffusivity relative to standard conditions.

where

u_e = migration velocity, cm/s
q_p = particle charge, coulomb
E = electric field strength, volts/cm

For a given field strength and particle charge, the effects of temperature and pressure are contained in the ratio "C'/μ_G" and are illustrated in Figure 1. The particle charge and electric field strength are complicated functions of temperature and pressure, and are beyond the scope of this study.

Gravitational Settling and Centrifugal Separation

Using equation (3) to describe the fluid resistance force, the gravitational settling velocity and the deposition velocity of a particle in a centrifugal force field may be approximated as:

$$u_s = \frac{1}{18} \frac{C' d_p^2 (\rho_p - \rho_G) g}{\mu_G} \tag{11}$$

$$u_c = \frac{1}{18} \frac{C' d_p^2 (\rho_p - \rho_G) u_t^2}{\mu_G R} \tag{12}$$

where

u_s = gravitational settling velocity, cm/s
u_c = centrifugal force deposition velocity, cm/s
g = acceleration of gravity, cm/s^2
u_t = tangential particle velocity at radius "R", cm/s
R = radial position of particle, cm
ρ_G = density of the gas, g/cm^3

In general, even at relatively high pressures ($<\sim 50$ atm), the gas density is much smaller than the particle density and may be neglected in equations (11) and (12). Therefore, the temperature and pressure dependence of equations (11) and (12) is contained in the ratio "C'/μ_G" and is illustrated in Figure 1.

Thermophoresis and Diffusiophoresis

Temperature and concentration gradients can give rise to deposition forces which can improve the collection efficiency of particulate control devices.

Thermophoresis is the result of gas molecules impinging on the particle surface from opposite sides with different mean velocities. The particle receives a net impulse opposite to the temperature gradient in the gas. The magnitude of the thermophoretic force was first devised by Epstein[5], and may be used with equation (1) for the resistance force, to obtain the thermophoretic deposition velocity as:

$$u_d = -\frac{3\,C'\mu_G}{2\,\rho_G\,T}\left[\frac{k_G}{2k_G + k_p}\right]\nabla T \qquad (13)$$

where

k_G = gas thermal conductivity, J/s-m-°K
k_p = particle thermal conductivity, J/s-m-°K
∇T = temperature gradient, °K/cm
u_d = thermophoretic deposition velocity, cm/s

Figure 5 shows the effects of temperature and pressure on the thermophoretic deposition velocity per unit temperature gradient, for a 5 μm diameter silica particle, as calculated from equation (13). The deposition velocity increases slightly with increasing temperature but decreases significantly with pressure.

More elaborate equations for predicting the thermophoretic deposition velocity have been presented by Hidy and Brock [6] and Derjaquin and Yalamov[7]. These equations are generally a function of the thermal-accomodation coefficient between the molecule and the particle surface. The accomodation coefficient has been shown to be a strong function of temperature (Byers and Calvert[8]), at relatively low temperatures. No information has been found con-

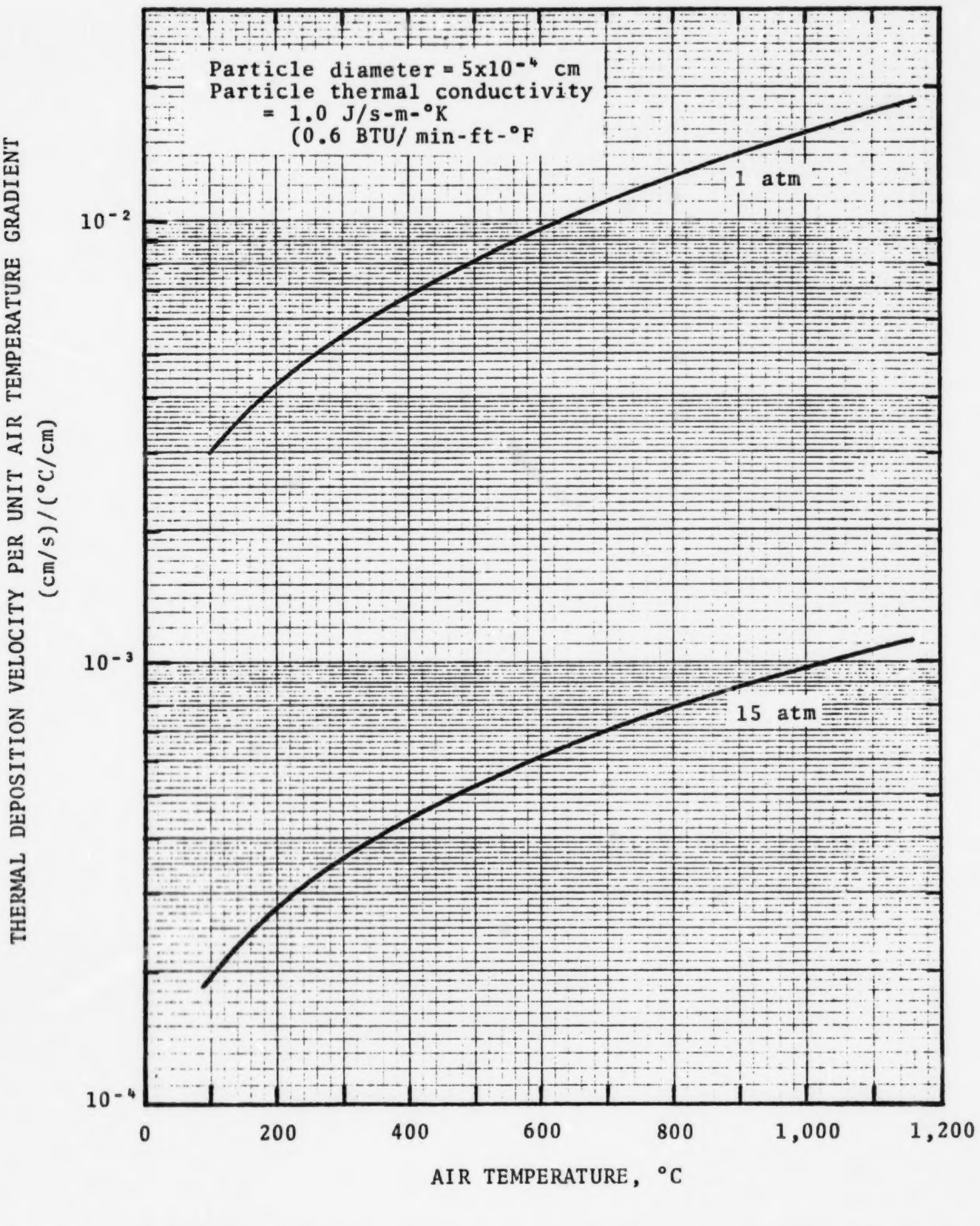

Figure 5. Thermal deposition velocity as a function of temperature and pressure

cerning the effects of very high temperatures and pressures on the accomodation coefficient. This lack of information represents a major uncertainty in the extension of the theory of thermophoresis to high temperature and pressure conditions.

Diffusiophoresis is generally important in situations where there are large concentration gradients, or where vapor condensation is occurring. It is unlikely that such conditions will exist in high temperature and pressure particulate removal systems. However, diffusiophoresis may be important in situations where high pressure and low or moderate temperature particle collection is required.

Calvert, et al.[3] looked at the problem of diffusiophoresis at high pressure and relatively low temperature. Figure 6 (from Calvert el al.[3]) shows the diffusiophoretic velocity as a function of vapor pressure gradient for various temperature and pressure conditions. The diffusiophoretic velocity is greatly reduced at high pressures.

Particle Size Alteration

One way to improve the collection efficiency for fine particles is to cause the fine particles to agglomerate into large aggregates which can be collected more easily.

Thermal coagulation - Thermal coagulation is the agglomeration of particles undergoing random Brownian motion. The rate of agglomeration (or coagulation) is generally considered to be proportional to the square of the particle number concentration. That is,

$$\frac{dN}{dt} = - K_o N^2 \qquad (14)$$

where

N = particle number concentration, cm^{-2}

t = time, s

K_o = coagulation constant, cm^3/s

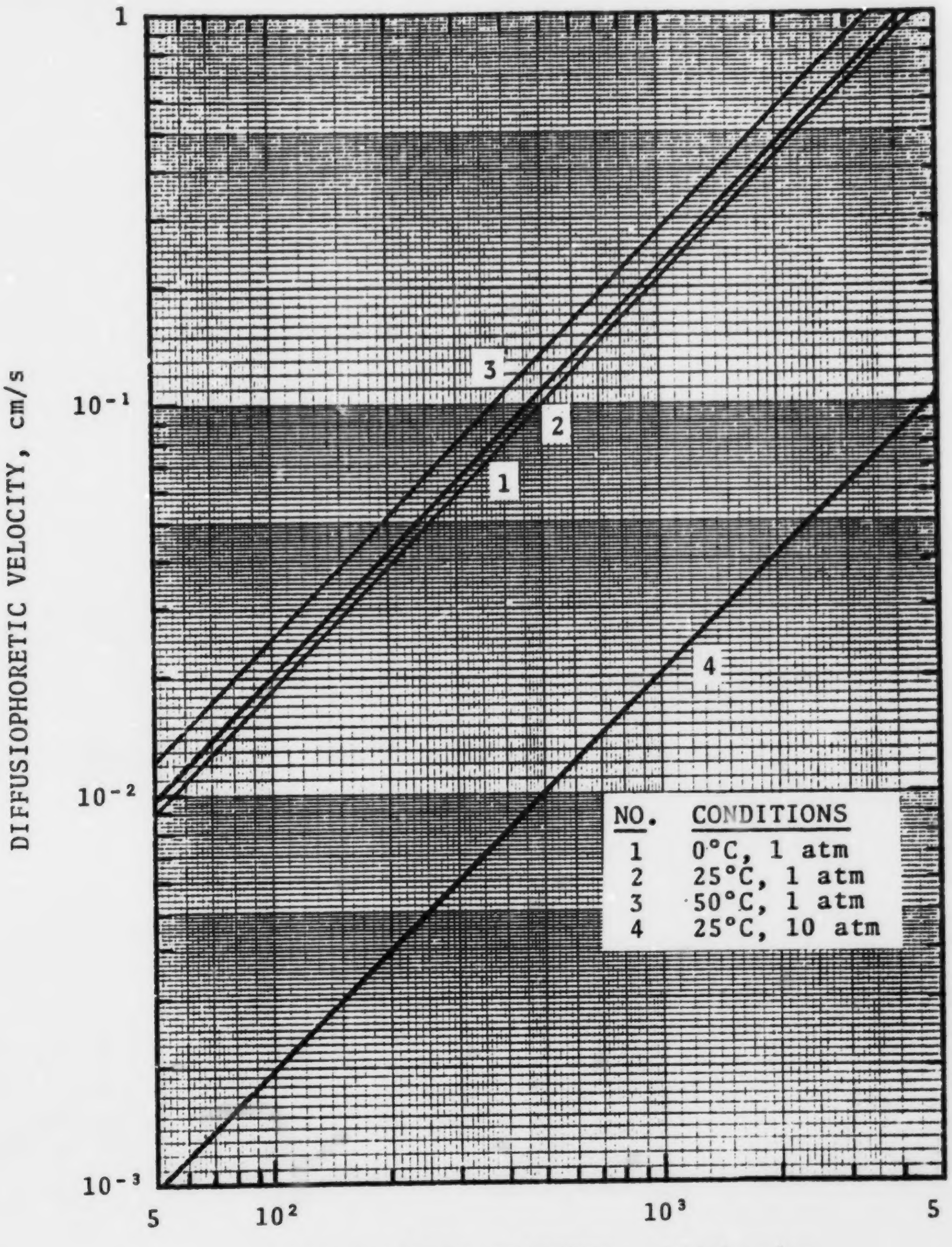

Figure 6. The effect of high temperature and moderate pressures on the diffusiophoresis of particles.

Using equation (3) for the fluid resistance, Fuchs[9] presents the following equation for the thermal coagulation constant.

$$K_o = 4\pi\ D\ d_p = \frac{4}{3}\ \frac{k\ T\ C'}{\mu_G} \qquad (15)$$

The thermal coagulation constant is shown as a function of temperature, pressure, and particle diameter in Figure 7. The agglomeration of particles increases with temperature and decreases with pressure. It appears that at high pressure and high temperature, there is a small increase in the rate of agglomeration for particles larger than 0.1 μm. At high temperature and atmospheric pressure, the rate of agglomeration of fine particles would be increased significantly, however, high pressures greatly decrease the benefit of high temperature.

Turbulent Coagulation - Particles can also agglomerate as a result of turbulence in the fluid. Turbulent coagulation has been discussed by Beal[10] and is proportional to the turbulent diffusion coefficient. For the general case where the particle diameter is much smaller than the turbulent microscale, Beal presents the following equation for the turbulent diffusion coefficient.

$$D_t = \frac{1}{4}\ \sqrt{\varepsilon_o/\nu}\ r_p{}^2 \qquad (16)$$

where

D_t = turbulent diffusion coefficient, cm^2/s

ε_o = energy dissipation rate for unit mass of fluid, cm^2/s^3

ν = kinematic viscosity, cm^2/s

r_p = particle radius, cm

Therefore, "D_t" is proportional to "$\nu^{-1/2}$". Figure 8 is a plot of "$\nu^{-1/2}$" against temperature for a range of pressure. Turbulent agglomeration increases greatly with an increase in pressure

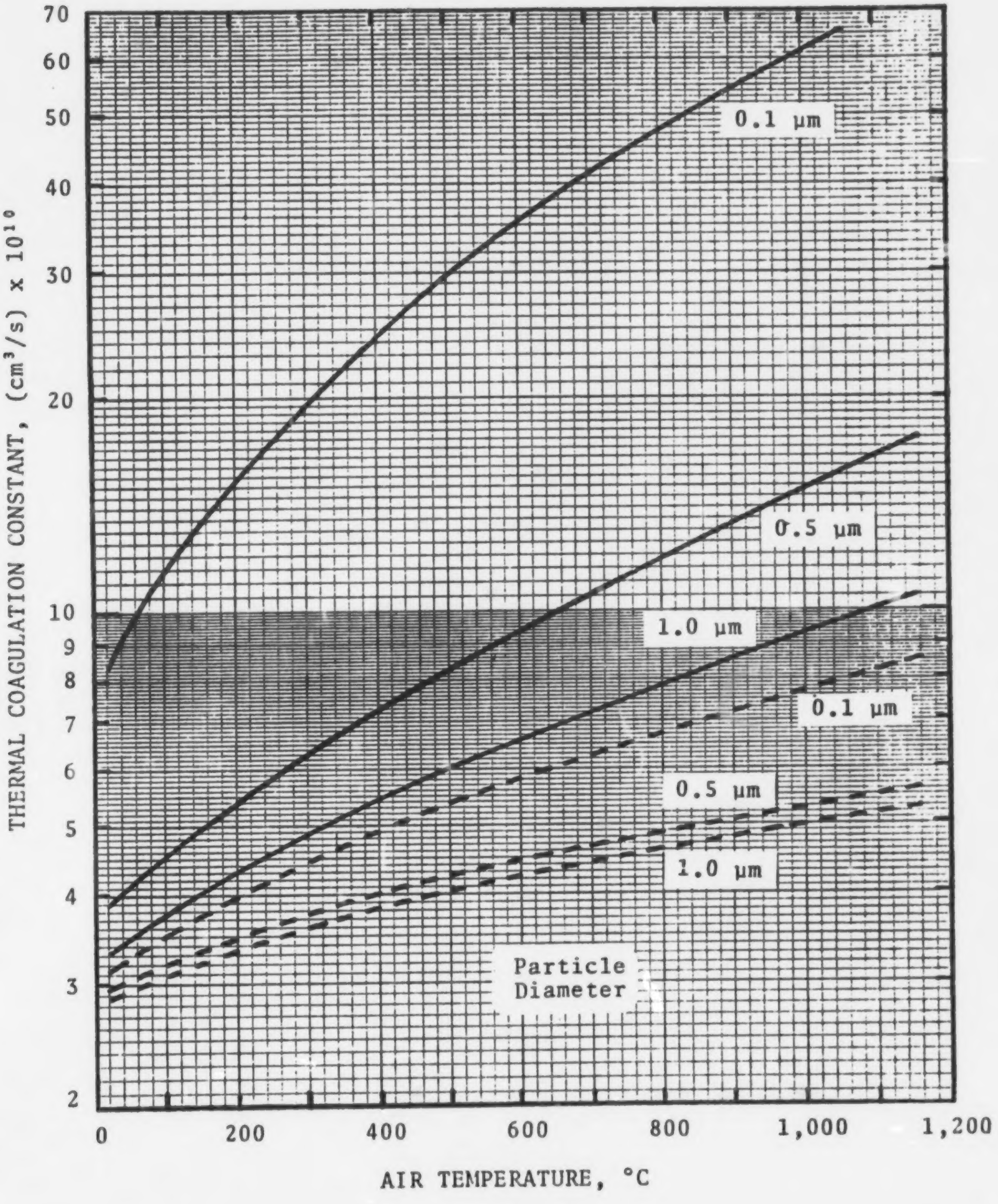

Figure 7. Thermal coagulation of particles at high temperature and pressure.

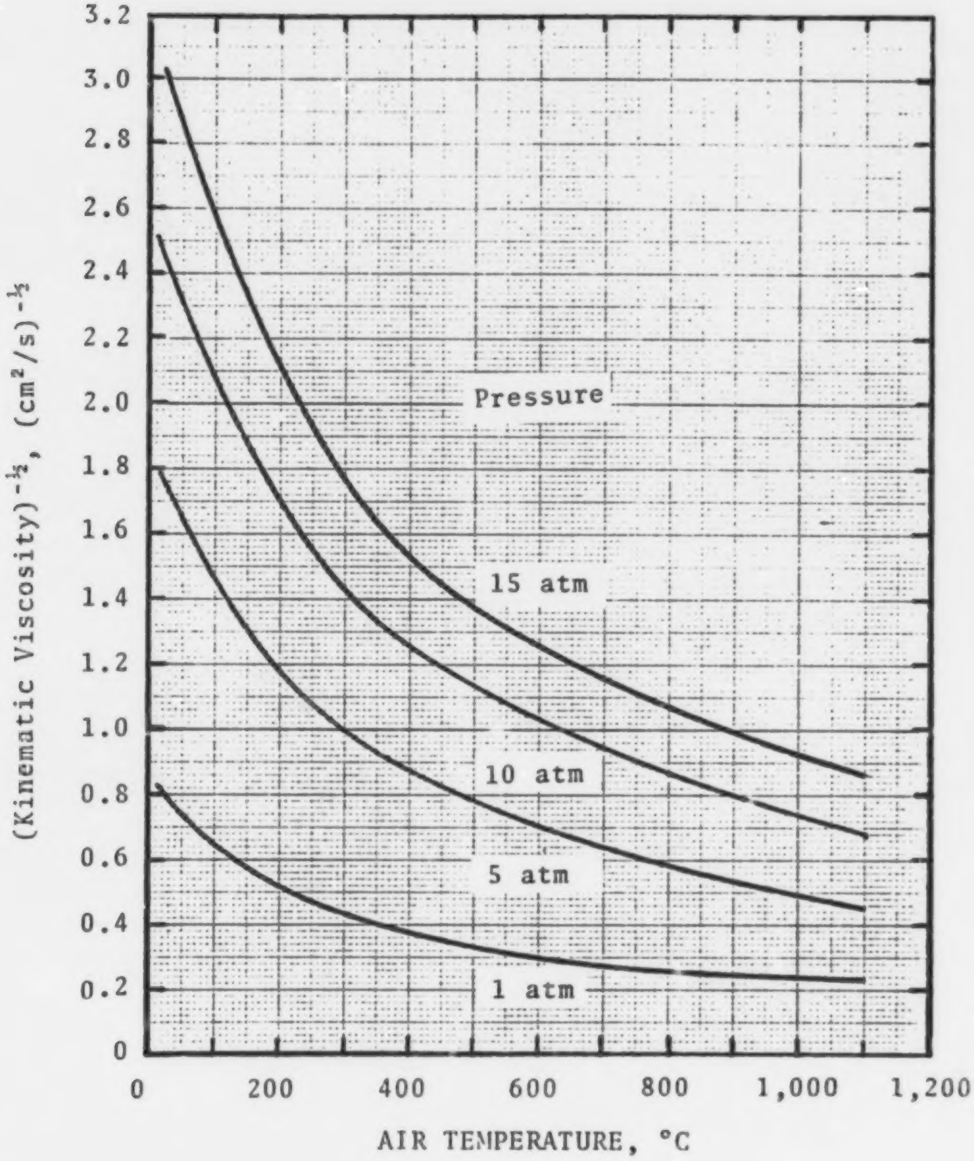

Figure 8. Turbulent agglomeration of particles at high temperature and pressure.

at low temperatures. Unfortunately this beneficial effect of pressure is almost completely nullified at high temperatures.

Sonic agglomeration - Another way to cause particles to agglomerate is by the application of sonic vibrations. Sonic agglomeration has been studied by Mednikov[11] and has been shown to be inversely proportional to the square root of the product of gas density and the speed of sound. That is,

$$K_a \sim \left(\rho_G \, C_G\right)^{-1/2} \qquad (17)$$

where

K_a = sonic agglomeration coefficient

C_G = speed of sound in the gas

The speed of sound of an ideal gas is given by:

$$C_G = (\gamma R T)^{1/2} \qquad (18)$$

where

γ = the ratio of specific heats, dimensionless

R = gas constant, erg/g-°K

T = absolute temperature, °K

Figure 9 shows the relative sonic agglomeration coefficient for a variety of temperatures and pressures. Once again, high temperatures can slightly improve agglomeration at atmospheric pressure, but high temperature and high pressure together significantly reduce sonic agglomeration relative to standard conditions.

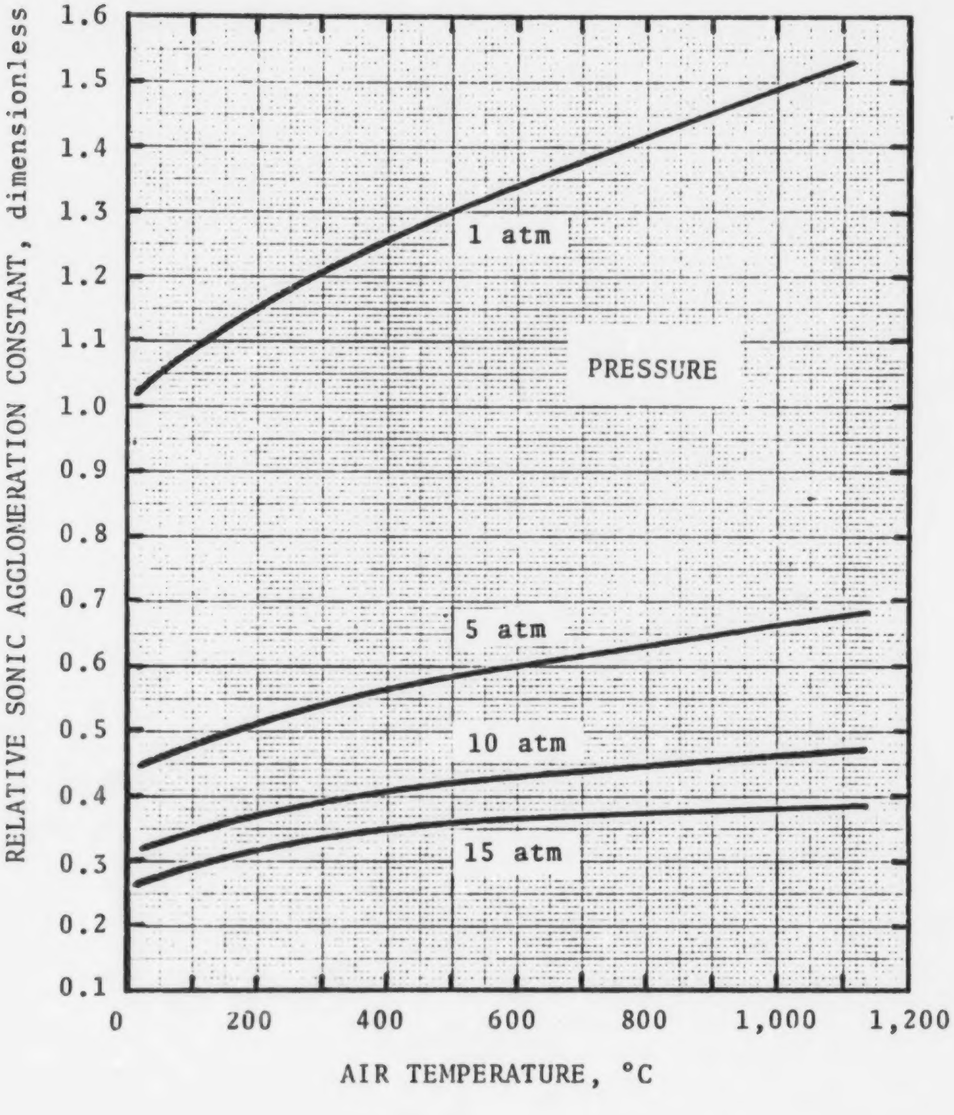

Figure 9. Sonic agglomeration of particles at high temperature and pressure.

EXAMPLES

Single Stage Impactor

Using equation (3) to represent the fluid resistance force, it is possible to predict the collection efficiency and work requirements for particle collection by inertial impaction. Figure 10 shows the particle cut diameter as a function of flow work (or specific power) for a single stage impactor at various temperature and pressure conditions. For an impactor, it may be assumed that all particles larger than the cut diameter will be collected. The flow work is equivalent to the energy requirement per mass of gas. The power requirement would be equal to the flow work multiplied by the mass flow rate.

Figure 10 illustrates that the work required to collect submicron particles by inertial impaction increases rapidly with decreasing particle diameter. The effects of temperature and pressure may be seen by comparing the four curves. Curve 1 represents standard temperature and pressure conditions. Curve 2 shows the effect of high temperature at atmospheric pressure. The work required to collect submicron particles is not increasing as rapidly as that required to collect larger particles. The reason for this is that the fluid resistance force decreases as the mean free path of the gas molecules increases relative to the particle diameter, and the mean free path increases with temperature, for a constant pressure.

Curves 3 and 4 are for simultaneous high pressure and high temperature conditions. The beneficial effect of high temperature on the mean free path is completely nullified by a decrease in mean free path with increasing pressure. The work required to remove submicron particles is greatly increased. For example, the collection of all particles greater than 1 μm by inertial impaction would require about 450 J/kg at standard conditions (Curve 1). It would require about 4,000 J/kg at 1,100°C and 15 atm, (Curve 4). This is equivalent to approximately a 9:1 increase

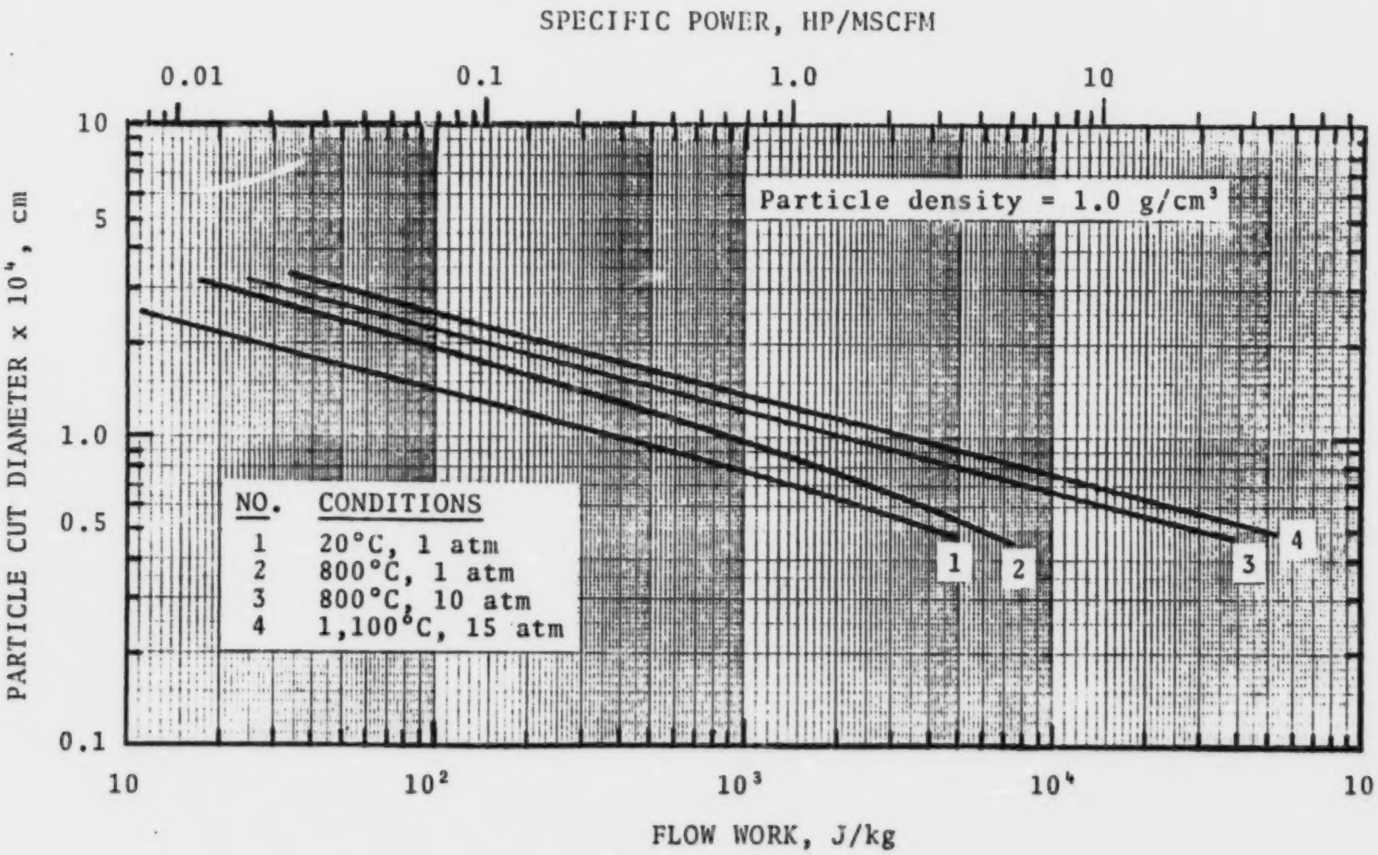

Figure 10. Flow work (specific power) for impaction from a round jet as a function of temperature and pressure.

in the power requirement to maintain a similar degree of particle removal.

Cyclone Separator Efficiency

A further illustration of the effect of high temperature and pressure on particle collection may be obtained by predicting the collection efficiency of a cyclone separator operating at various temperatures and pressure, for the same inlet velocity.

Figure 11 shows a typical efficiency curve for a high efficiency cyclone (curve 1). The cyclone cut point (50% efficiency) occurs at a particle diameter of about 1 μm, and the cyclone is better than 99% efficient for particles larger than 15 μm. This curve was obtained using the theory presented by Calvert, et al[3] and attributed to Leith and Licht[12]. The cyclone was assumed to be about 15 cm (6 inches) in diameter with a volumetric flow rate of about 1,400 ℓ/min (50 ft^3/min).

Curve 2 shows the estimated cyclone efficiency for a gas at 1,100°C and atmospheric pressure, and for the same inlet gas velocity as in curve 1. The cyclone efficiency has dropped significantly and now has a cut point occurring at 2.0 μm and is only 96% efficient for 15 μm particles.

Curve 3 shows the estimated cyclone efficiency for a gas at 1,100°C and 15 atm, for the same inlet velocity. The cyclone efficiency has decreased again slightly for small particles, but is relatively unaffected for larger particles. The cut point now occurs at 2.5 μm and the efficiency for 15 μm particles is 95%.

Specific Power for a Cyclone Separator

Figure 12 presents another example of high temperature and pressure effects on the performance of a cyclone separator. The specific power ratio is the ratio of the specific power (HP/MSCF) at high temperature and pressure to that at standard conditions. The curves in Figure 12 show the power requirement relative to standard conditions to collect various particle sizes while maintaining a constant collection efficiency for each particle size.

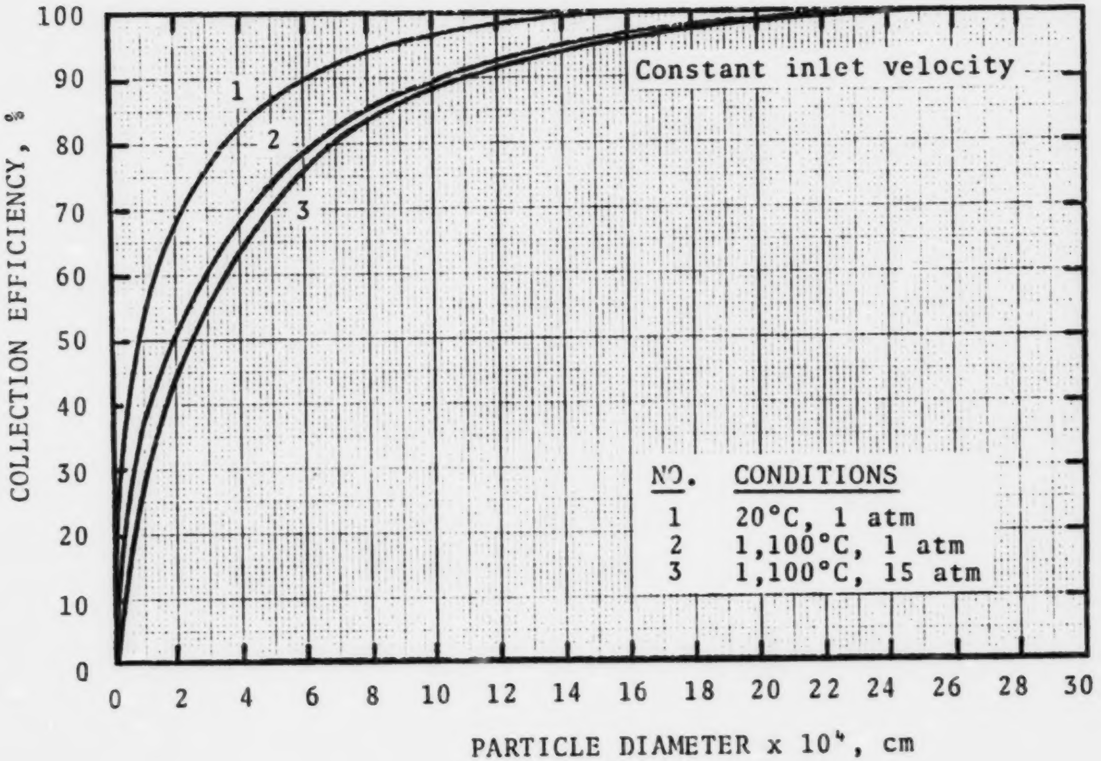

Figure 11. The effects of high temperature and pressure on the collection efficiency of a high efficiency cyclone.

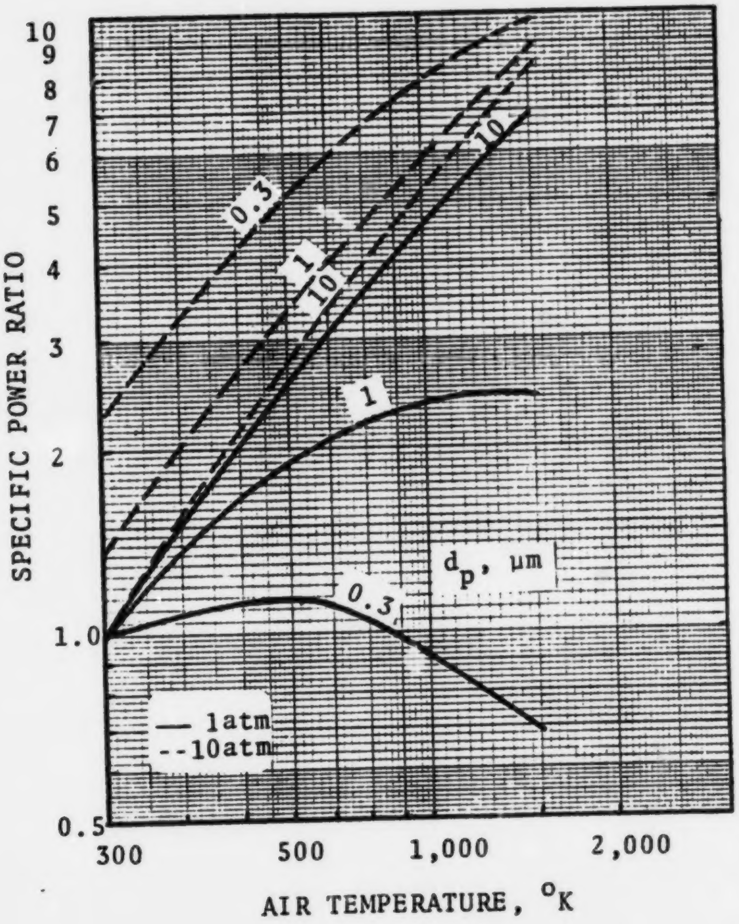

Figure 12, The specific power requirements for a cyclone as a function of temperature and pressure.

Efficiency of a Fiber Bed

Figure 13 shows the effects of temperature and pressure on the collection efficiency of a fiber bed. The curves were calculated from the theory presented by Calvert, et al.[3], and attributed to Jorgeson. The collection efficiency is the combined efficiency resulting from inertial impaction, interception, and Brownian diffusion.

From Figure 13 it is apparent that the filter collection efficiency for particles larger than about 0.5 μm in diameter is reduced significantly at high temperature and pressure. This is a result of the larger inertial impaction parameter (equation 2) at high temperature and pressure. For particles smaller than 0.5 μm, the collection efficiency is somewhat increased because of the increased Brownian motion at high temperatures.

At high temperature and atmospheric pressure the collection efficiency is greatly increased for particle diameters up to about 0.9 μm. For particles larger than 1 μm the collection efficiency is reduced at high temperature and atmospheric pressure, although not as severely as at high temperature and high pressure.

CONCLUSIONS

In general the particle collection efficiency is significantly reduced at high temperature and pressure for particles with diameters larger than a few tenths of a micrometer. Particle collection at high temperature and pressure is likely to be very difficult, especially for fine particles.

Some mechanisms seem to increase significantly at high temperature and low pressure (for example, Brownian diffusion). In this case, it is likely that collection of submicron particles would be easier at higher temperature. However, the collection of larger particles would still be more difficult.

It is unlikely that a new device will be devised which can remove particles from high temperature and pressure gases more efficiently than at standard conditions. It is more probable that

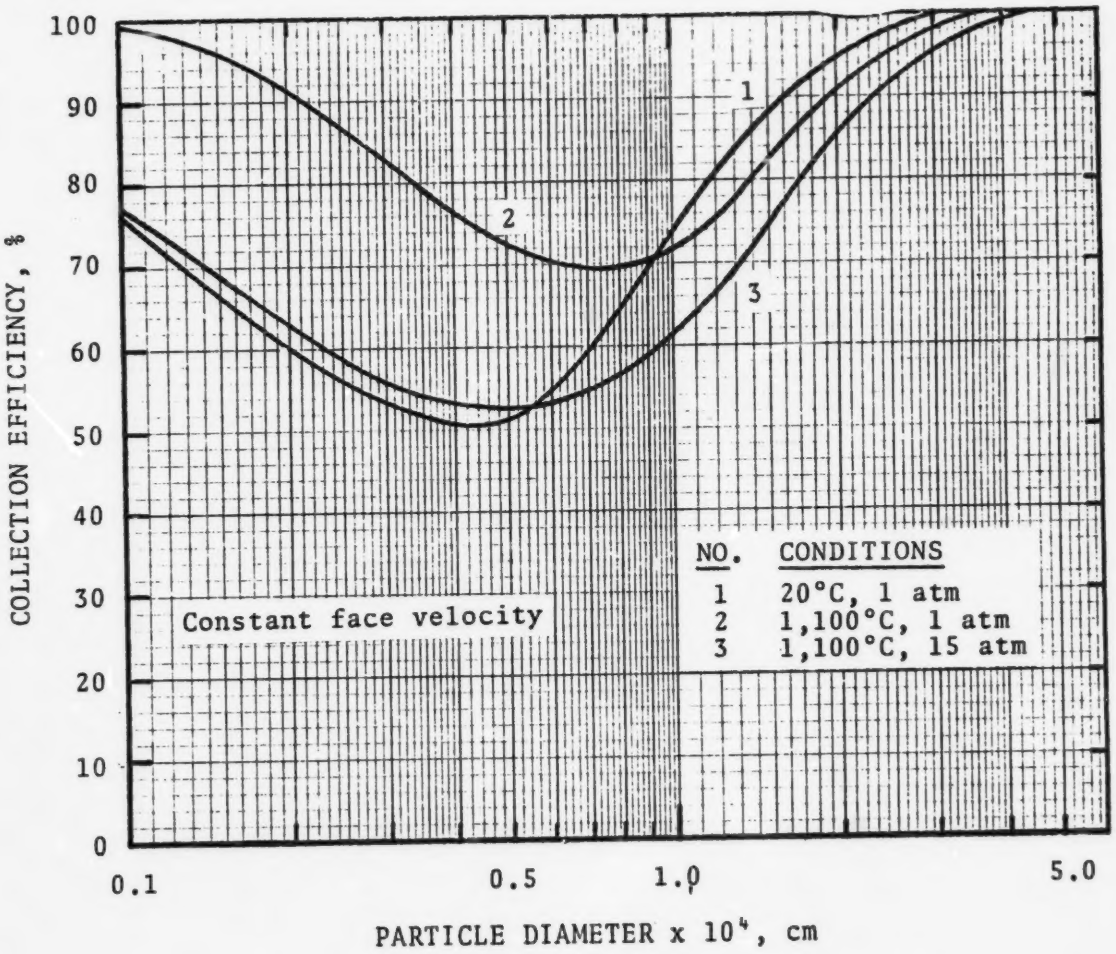

Figure 13. The effects of high temperature and pressure on the collection efficiency of a fiber bed.

high temperature and pressure particle cleanup will have to be achieved by modified conventional equipment (cyclones, metal or ceramic filters, granular beds) operating at higher costs (larger power consumption) than would be needed for operation at standard conditions.

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MEASUREMENT METHODS AT HIGH TEMPERATURE AND PRESSURE

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Symposium on

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ABSTRACT

The current evolution of coal conversion processes has generated a need for a new class of particulate and gaseous sampling instrumentation. Many new processes, such as fluidized bed combustion, coal gasification, and combined cycle power plants will operate at conditions of high temperature (some greater than 1000°C) and high pressure (up to 100 atmospheres). The combined risks of temperature, pressure, corrosion, and toxic and flammable chemicals will present a host of difficult problems in mechanical design, sample handling procedures, and safety when designing sampling systems for these processes.

This paper presents a brief review of some approaches which have been, or are being, used to sample HTHP processes, and a discussion of some of the major problem areas which must be considered in HTHP sampling.

MEASUREMENT METHODS AT HIGH TEMPERATURE AND PRESSURE

INTRODUCTION

A very significant obstacle facing the developers of coal conversion technology today is a lack of information on the service conditions of the conversion processes themselves. For many processes the actual gas composition, particle loading and physical state are not known with any significant degree of certainty. What data are known are based on bench-scale experiments or computer simulation. Getting this information, from pilot- or full-scale facilities, is required for the proper design, process control, and pollution abatement of these processes.

However, many of the new coal conversion processes will operate at conditions of high pressure (up to 100 atmospheres) and high temperatures (some greater than 1000°C). Sampling at these conditions will require the development of a whole new class of sampling technology. The combined risks of temperature, pressure, corrosion, and toxic and flammable chemicals present a host of difficult problems in mechanical and thermal design, sample handling procedures, and safety when designing sampling hardware for these conditions.

The hardware, procedures, and methodologies to accomplish these HTHP sampling tasks do not yet exist in any standard form or approach. It is not the purpose of this paper to solve these problems or even suggest any standard approaches. Only research and development will do that. We will, however, indicate some of the approaches that have been used and are being used (a cursory state-of-the-art survey) and identify the major problems which HTHP sampling will pose.

STATE-OF-THE-ART IN HTHP SAMPLING

State of the art in HTHP sampling technology is something of a mixed bag of tricks. In general, HTHP sampling environments have not lent themselves to standard sampling approaches; each specific process tends to have its own set of difficulties. Examples of HTHP hardware one finds in the literature are generally custom designed to solve specific access or technical problems, sometimes with strict economic constraints as well. An exception to this is in the area of flame composition sampling where there are commercial suppliers who offer off-the-shelf and specialty items as standard product lines (see Figure 1).

Space does not permit a comprehensive state-of-the-art review of HTHP sampling techniques, but the following paragraphs present some notable approaches that have been or are being used.

Sampling in the Bureau of Mines Coal-Fired Gas Turbine

In the early 1960's the Bureau of Mines conducted extensive development work on a coal burning gas turbine at its Morgantown (W. Va.) Coal Research Center. Coal ash erosion of the turbine blades was a critical factor, the evaluation of which required accurate and reliable knowledge of the duct characteristics at the turbine inlet and exhaust. A. S. Moore of the USBM developed a rather sophisticated particulate sampling probe for sampling the 19-inch I.D. ducts leading to the blade sections.(2)

Gas temperature and pressure for this application were 1300°F to 1500°F and 70 to 80 psig respectively. The probe was 1/2-inch I.D., 316 S.S., with nozzles of either titanium carbide or Hasteloy HS-31. Both the probe and the protube were water cooled. The probe had a powered traverse mechanism for profiling the duct (Figure 2(a)).

Australian Coal Burning Gas Turbine Research

The Australians have also conducted extensive direct-fired coal combustion research between the years 1948 to 1970. Much of their work was directed at determining the effects of blade erosion due to particulates and to characterize such erosion it was necessary to sample both mass and size of the particulate entering and exiting their turbine.

They obtained their data by direct sampling and counting particles. Particulates were collected either by small cyclones or absolute type filters.(3) More detailed data is not available at this time.

Gas Sampling at the Bureau of Mines Stirred-Bed Reactor (Gasifier)

Figure 2(b) shows a schematic of the gas sampling and analysis system also used at the Morgantown Coal Research Center for sampling their stirred-bed gasification process currently under development.

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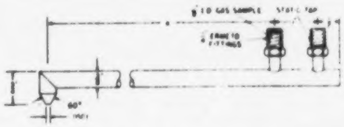
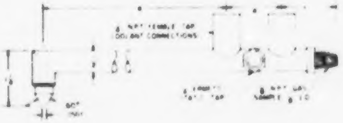
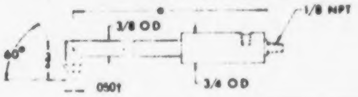
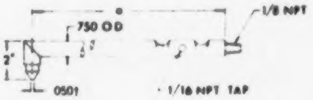
SELECT FROM THESE STANDARD GAS SAMPLING TYPES AND MODELS

Description	Ordering Part No.	Probe* Length†	Dimensions
TYPE GA Uncooled, of welded construction, this probe is suitable for use up to 2000°F and is the smallest standardly offered.	GA-12-.050 GA-24-.050	12" 24"	
TYPE GB (Inconel Tip) This probe can be used either as a heated or cooled probe. The cooling passages are especially designed for unrestricted flow of cooling or heating fluids. For some applications, it is important to keep the sampled gas above 500°F to the analyzer to prevent condensation of evaporated hydrocarbons. In this case, the probe is heated by blowing hot air through the cooling passages. This is more satisfactory than trying to heat it with an electric coil because the variation of heat loss conditions from the surface of the probe makes it almost impossible to maintain an even temperature. For higher temperatures, the probe can be kept hot enough to prevent condensation but cool enough not to melt by blowing unheated shop air through the coolant passages. Type GC is recommended for temperatures exceeding 3000°F.	GB-12-.050 GB-24-.050	12" 24"	
TYPE GC (Copper Tip) This probe has a copper tip with high enough conductivity to prevent burning out at the higher temperatures and velocities. Since the tip is easily damaged, it is not recommended for lower temperature applications.	GC-12-.050 GC-24-.050	12" 24"	
TYPE GA-F (Internal Pressure & Water Taps) All Stainless Steel construction. Reinforcement at pressure take off section offers female tap for ease of installation. Performs exactly like Type GA.	GA-F-12-.050 GA-F-24-.050	12" 24"	
TYPE GB-F (Internal Pressure & Water Taps, S/S Tip) Same as Type GB except internal pressure taps. Barrel is ¾" O.D. and sensing head is ½" O.D. Specially designed for ease of installation.	GB-F-12-.050 GB-F-24-.050	12" 24"	
TYPE GC-F (Internal Pressure & Water Taps, Copper Tip) Same as Type GC except internal pressure taps. Barrel is ¾" O.D. and sensing head is ½" O.D.	GC-F-12-.050 GC-F-24-.050	12" 24"	

*Longer or shorter lengths available. Consult United Sensor.
†Inlet holes may be enlarged upon request.

Figure 1   Standard gas sampling probes.(1)

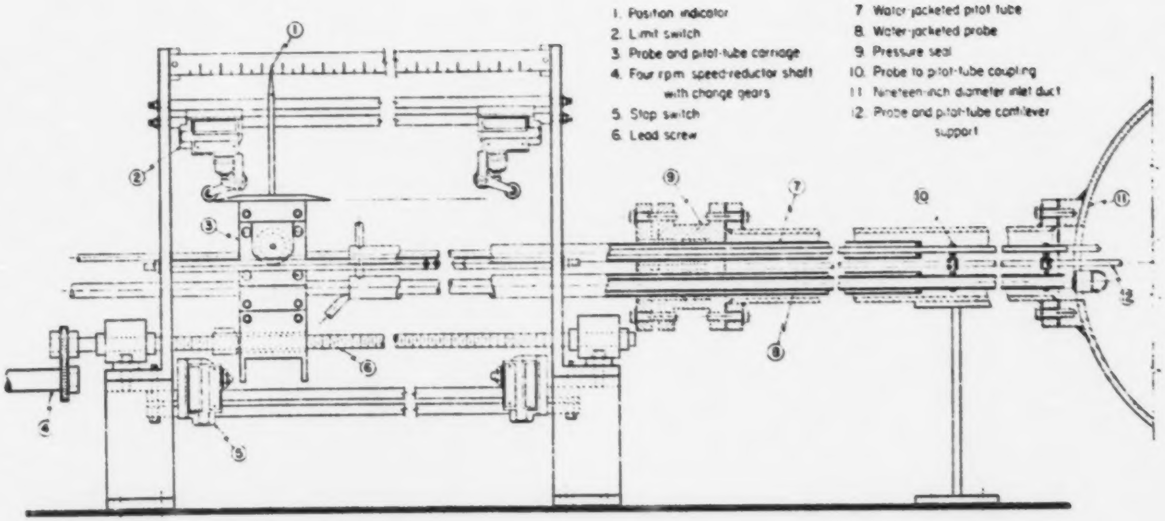

1. Position indicator
2. Limit switch
3. Probe and pitot-tube carriage
4. Four rpm speed-reductor shaft with change gears
5. Stop switch
6. Lead screw
7. Water jacketed pitot tube
8. Water-jacketed probe
9. Pressure seal
10. Probe to pitot-tube coupling
11. Nineteen-inch diameter inlet duct
12. Probe and pitot-tube cantilever support
①
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Figure 2(a) Sampler for USBM coal burning gas turbine.[2]

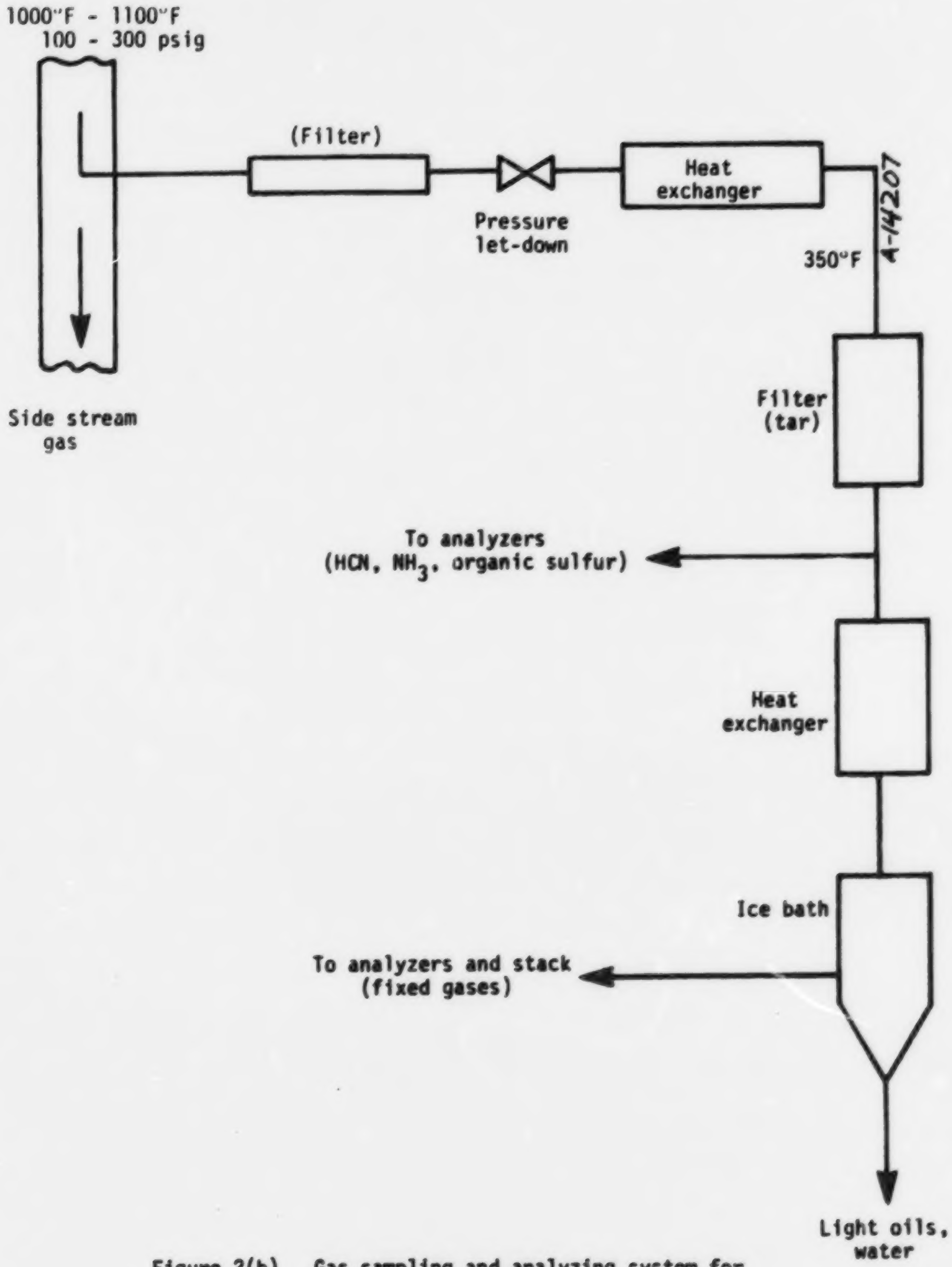


Figure 2(b) Gas sampling and analyzing system for minor constituents.

Stream conditions are 100 to 300 psig, 1000°F to 1100°F. Particulates are caught at high temperature in a bulk filter. After pressure and temperature reduction, the condensing tars are caught on an alundum or silicate filter; gases are analyzed at several stages in the sampling process, as shown.

Sampling at the BCURA Fluidized Bed Installation, Leatherhead, England

Recent telecons with personnel at the BCURA pressurized fluidized bed combustor in England have revealed how they approached sampling their process. Figure 3 is a drawing of their pilot-scale MKII combustor; it substantially represents the real installation.(4)

BCURA sampled for particulate size at 800°F to 925°F and 4 atmospheres following their primary and secondary cyclones on the horizontal run exiting the second cyclone, which is a 10-inch I.D., 4-foot long duct followed by a 6-inch I.D., 4-foot long water quench section.

Access to the duct was through a gland installed on an elbow at the end of the water quench section. They inserted a 5/8-inch S.S. probe axially some 4.5 feet up the horizontal duct (which gets the sampling nozzle ahead of the quench section) and withdrew a sample isokinetically (120 ft/sec), collecting samples in a cyclone. Sizing was done by coulter counter. Particles have been found to be primarily less than 5 microns.

Westinghouse Developed Sampler for Pressurized Fluidized Bed Combustor Sampling

Also in the area of pressurized FBC process sampling, Westinghouse engineers designed and constructed a small single point sampling system which consisted of a sampling probe (~1/4-inch tubing), static pressure probe, impactor (Brinks or Anderson) scalping cyclone, heated oven, and miscellaneous valves and hardware. Figure 4 presents the system somewhat schematically.

The device was designed to be installed and removed only when the process was not in operation. The impactor was placed within a pressure chamber inside the oven (350° capability) so that neither it nor the oven had to withstand system total pressure (10 atmospheres).

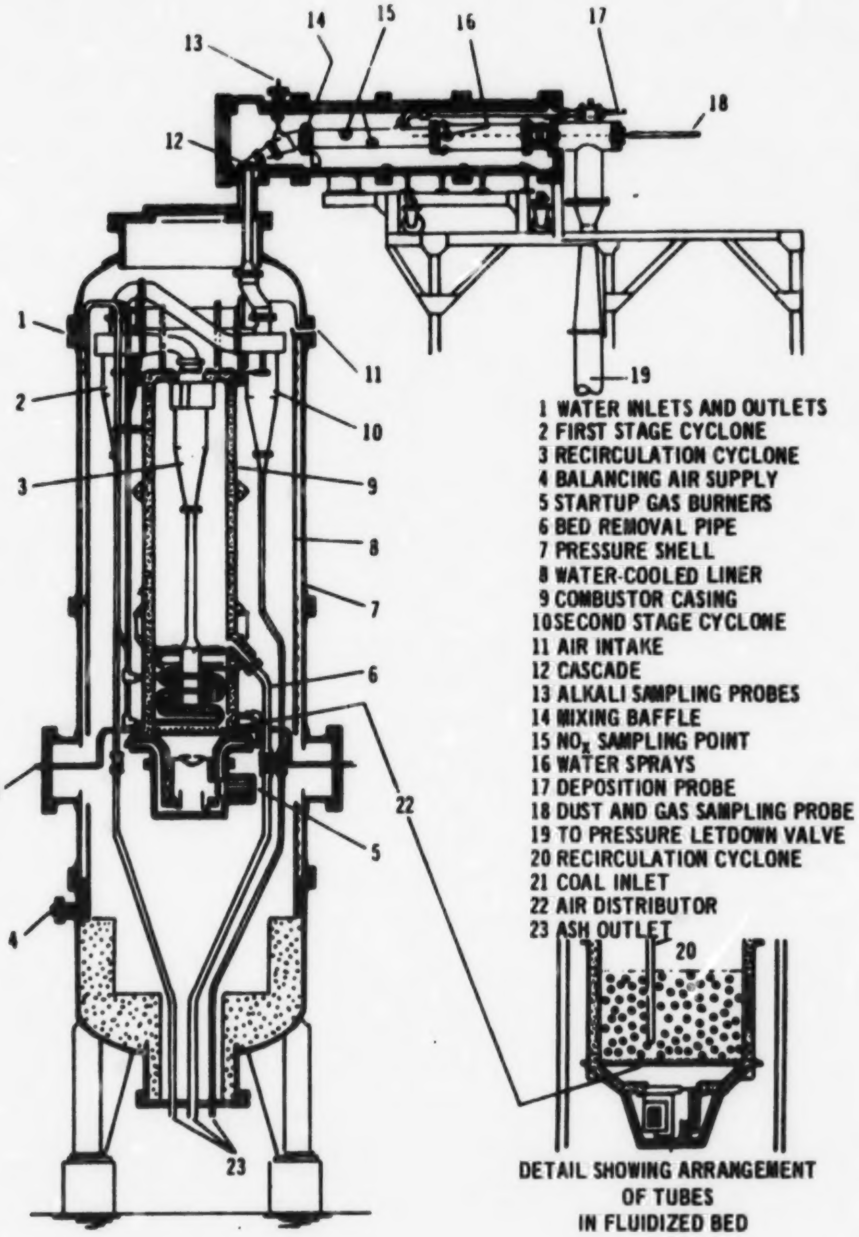


Figure 3 BCURA pressurized fluidized bed combustor (Reference 4).

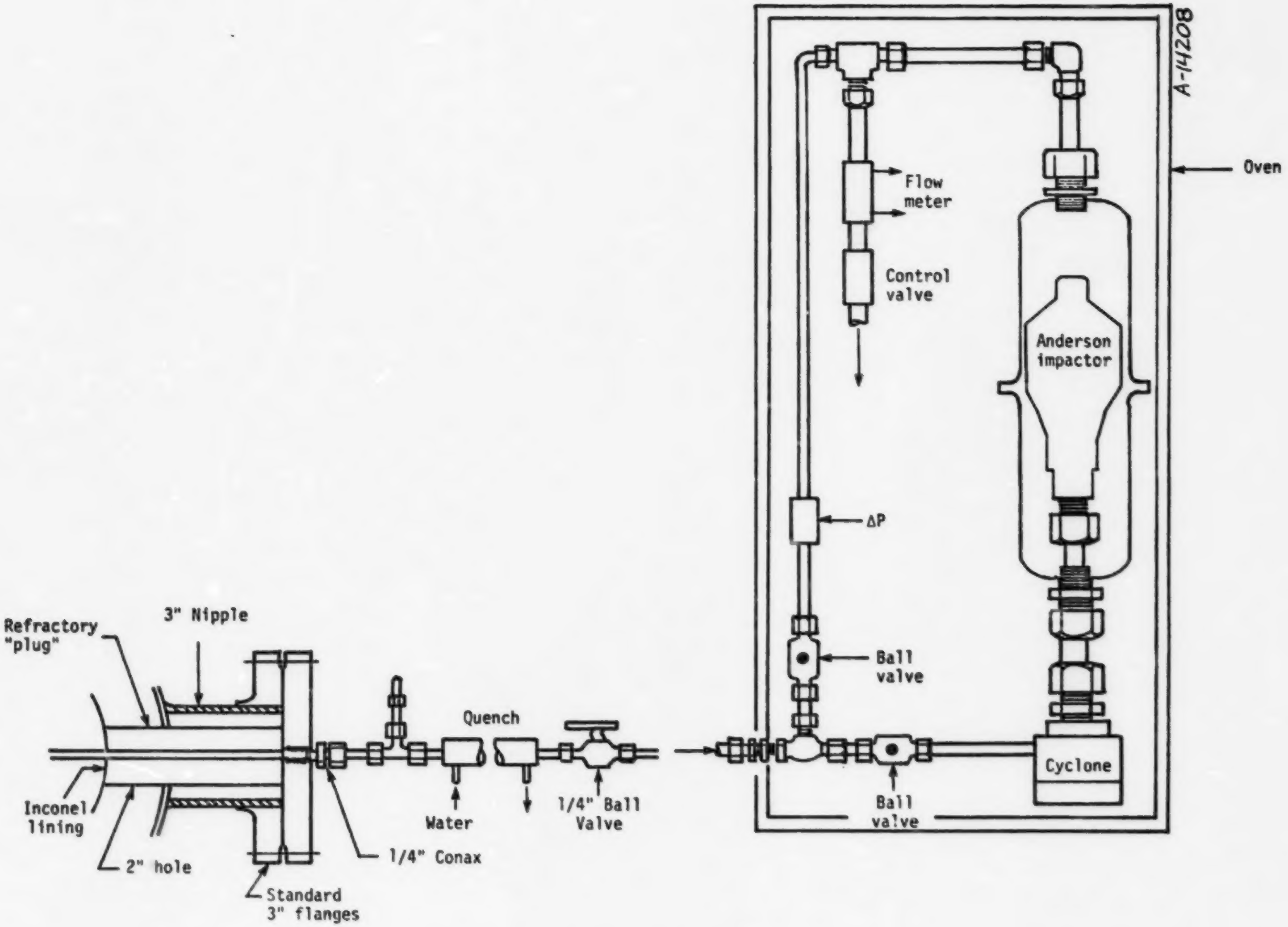


Figure 4 Westinghouse 10 atm particulate sampler.

High Temperature, High Pressure Sampling at Aerotherm

Aerotherm Division of Acurex Corporation has developed several sampling probes in the past 10 years which have been applied to high temperature and/or high pressure environments.

Figure 5 shows a miniature water-cooled gas sampling probe utilized in pinpoint sampling of high enthalpy, high velocity gas flows typical of those issuing from an arc-plasma generator. Environmental conditions included temperatures to 7000°F, 20 atmospheres total pressure, 15,000 ft/sec velocities and corrosive gases. Survival in this extreme environment was attained by a high performance liquid cooling system which employed nucleate boiling phenomena to affect adequate heat transfer rates. Coolant flow pressure drops were on the order of 1000 psi. A similar probe has been supplied to the Combustion Research Section of the EPA for flame sampling.

A somewhat more conventional example is shown in Figure 6 which shows one of a line of water-cooled stack sampling probes designed for use at temperatures up to 2400°F. The nozzles, which were uncooled, were fabricated of Inconel ^{R} for high temperature resistance. In some applications, special high temperature Cermet coatings such as Sermatel ^{R} were applied to provide corrosion resistance.

During the past year, Aerotherm has nearly completed development of a sophisticated particulate train for the Navy designed for operation in the exit plane of military jet engines. The system, shown in Figures 7 and 8, will automatically isokinetically sample the jet stream over a velocity range of Mach 0.1 to Mach 1.3. Gas stream conditions include temperatures to 1400°F and pressures to 50 psi. The system gives real-time particle size distribution data for submicron particles by means of a charge mobility analysis device by Thermo-Systems, Inc.

A rather interesting feature of this system is a microprocessor with programming capability which permits the rapid automatic calculation and flow control necessary to respond to engine operating point changes, and which will ultimately permit automatic positioning of the probe anywhere on a two-dimensional x-y plane.

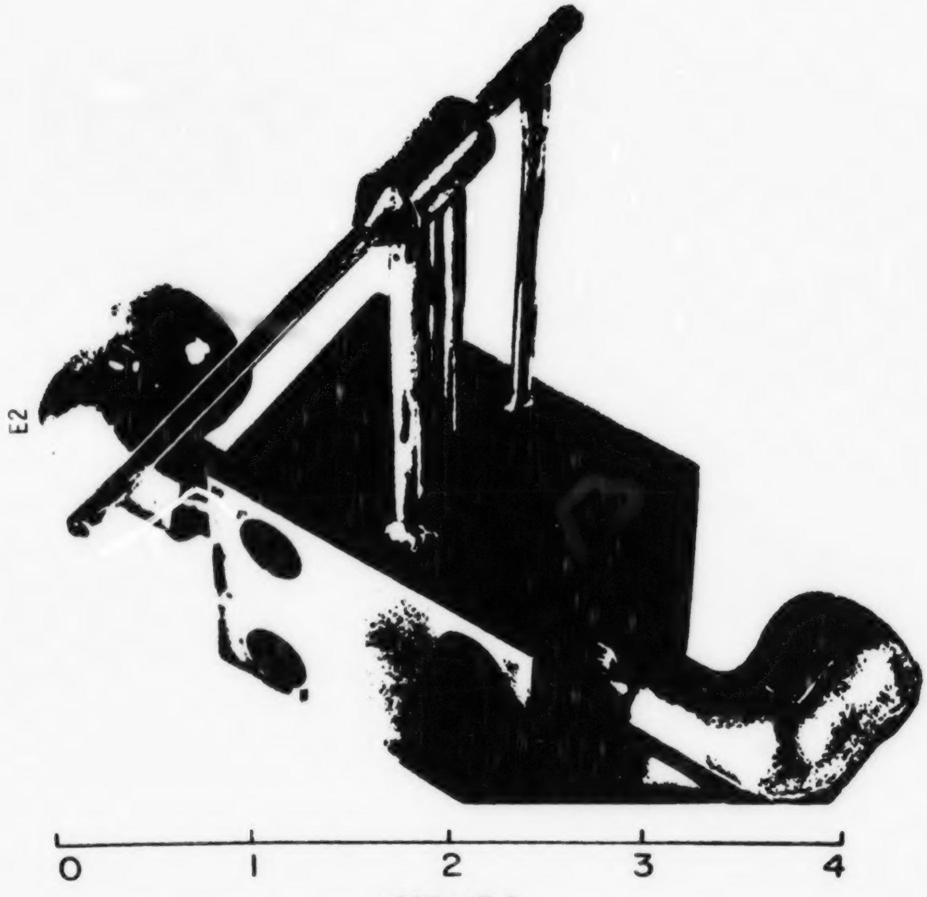


Figure 5 Aerotherm Miniature Water-Cooled Gas Sampling Probe

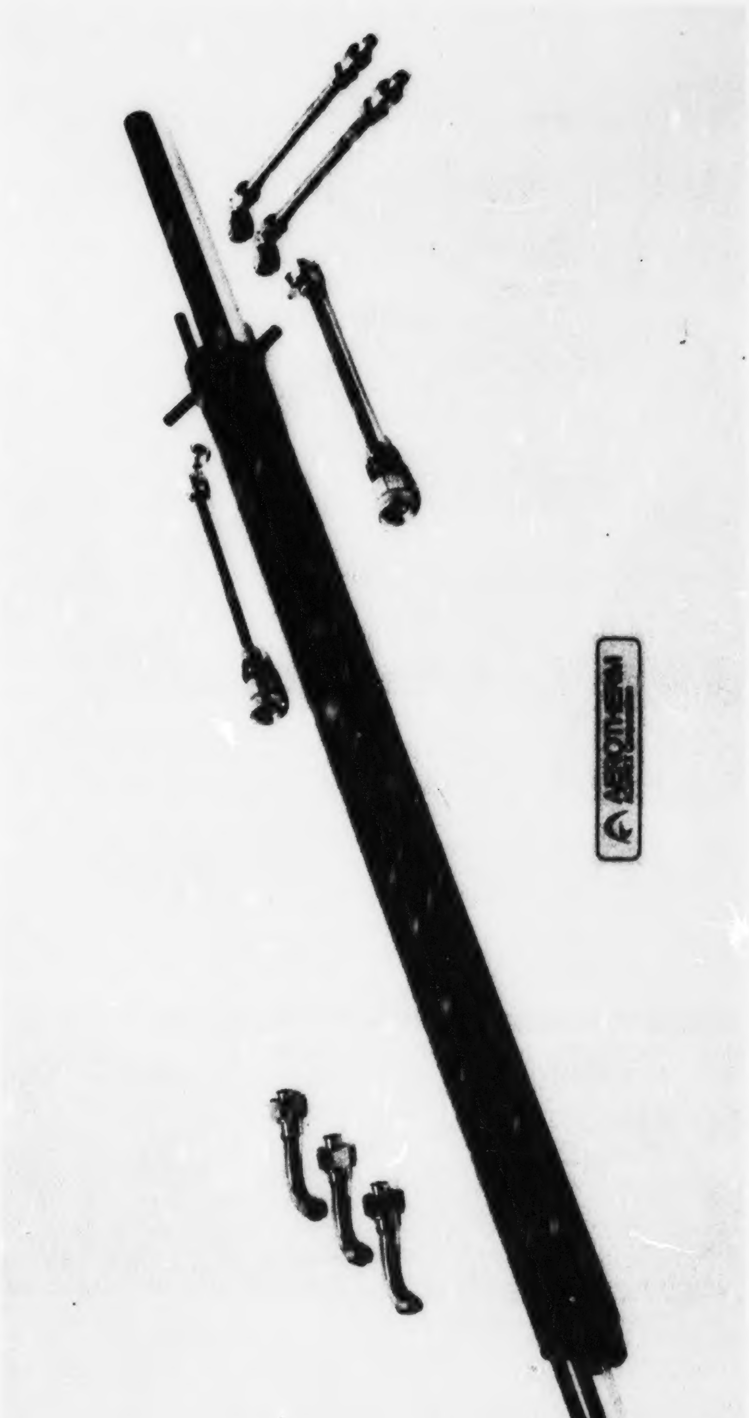
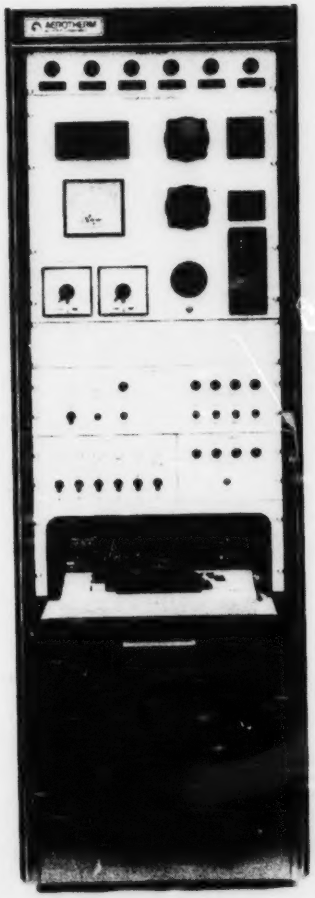
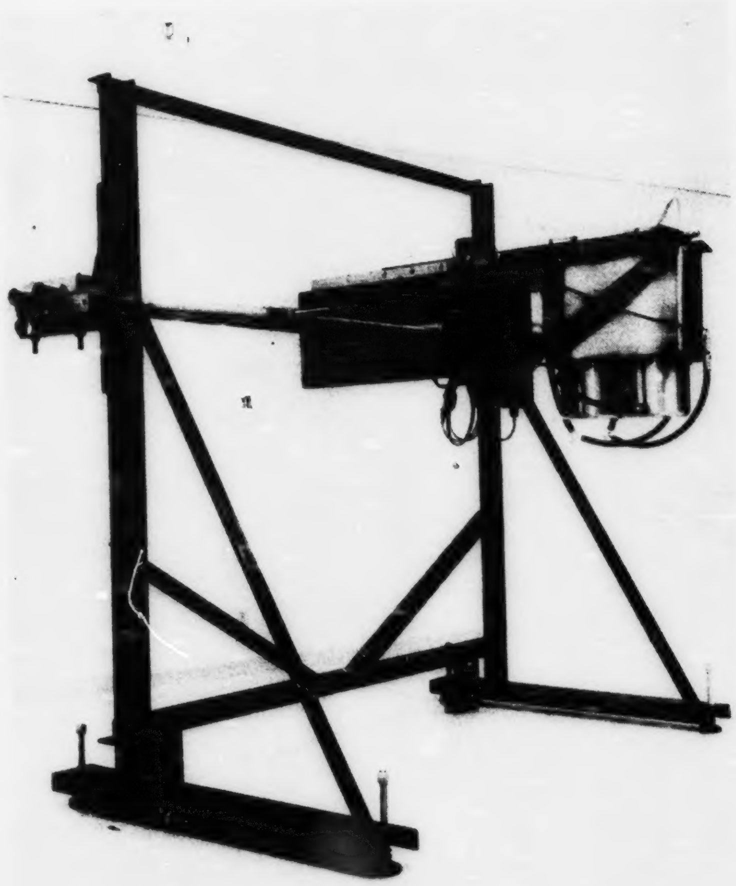


Figure 6 Water cooled probe.





Jet engine sampler.

Figure 7 Control console with microprocessor. Figure 8 Sampling train and support stand.

Current HTHP Sampling Research and Development at Aerotherm

Aerotherm currently has under development (for the EPA), a sampling system designed to acquire particulate mass loading and size distribution data in process gas streams such as pressurized fluidized combustion processes. The system was designed to be applicable to a rather broad range of process conditions, shown in Table 1. As a result it contains certain features not normally associated with process-specific systems such as those we have been reviewing in the foregoing paragraphs.

The probe is hydraulically activated, due to the high blow-off loads experienced at process pressures up to 20 atmospheres. Probe cooling and sample conditioning is accomplished by a Dowtherm circuit. Dowtherm permits sample conditioning to 600°F regardless of process conditions and avoids the "cold wall" condensation problem. The specified temperature and pressure range represents a turn-down of approximately 10:1 in sample mass flowrate. Figure 9 presents a pictorial overview of the system; several detail views of the probe assembly are presented in Figure 10.

Development of this system is being carried out as one subtask of an extensive level-of-effort program sponsored by the Industrial Environmental Research Laboratory (W. Kuykendahl, Project Officer) entitled "Measurements for High Temperature, High Pressure Processes," The total program extends 3 years and is basically structured to extend the state of the art in particulates and gaseous sampling technology to 1500 psi, 2000°F environments. Emphasis will be on coal conversion processes (primarily gasification) and pressurized fluidized bed combustion. Program structure is shown in Figure 11.

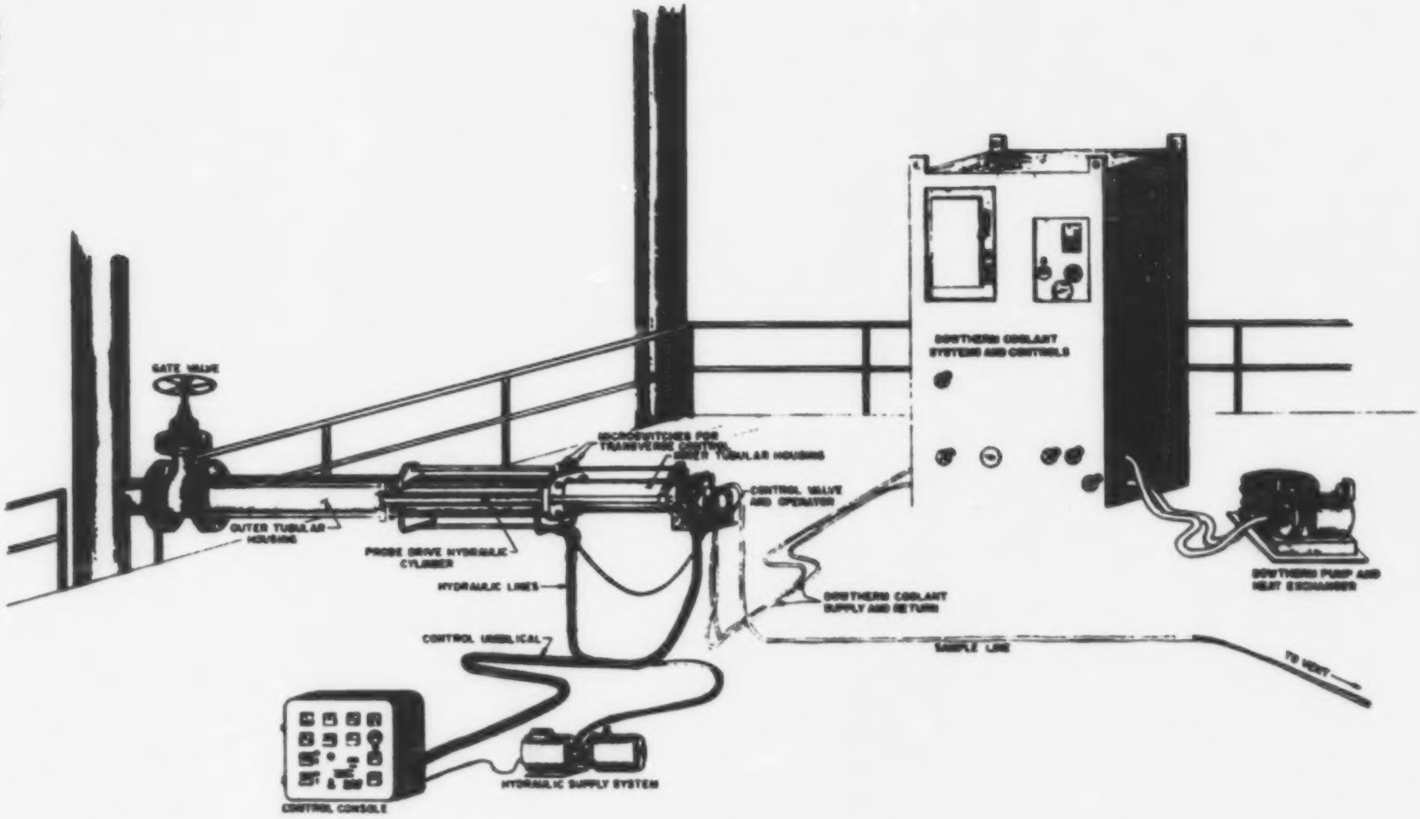
MAJOR PROBLEM AREAS ENCOUNTERED IN HTHP SAMPLING

High temperature, high pressure sampling presents several formidable problem areas to the system designer:

- Pressure and temperature
- Corrosion and erosion
- Condensation of process constituents
- Toxicity of process constituents
- Flammability and explosion hazard of process constituents

TABLE 1 HIGH TEMPERATURE, HIGH PRESSURE SAMPLING SYSTEM SPECIFICATIONS

Sample Environment	
• Temperature	1200°F - 1800°F 650°C - 1000°C
• Pressure	3 - 20 atmospheres
• Stream Velocity	15 - 100 fps
• Particulate Grain Loading	0 - 15 gr/ft^3
• Particulate Size Range (for classification)	0.2 - 7.0 microns
• Duct Size	Variable depending on probe; std. is 8 inches I.D. minimum
Traverse Capability or Penetration of Nozzle into Duct or Vessel	Approximately 26 inches either in situ or extractive configuration (some dependence on internal configuration of duct vessel)
Access Process Port Requirements	Standard: 4 inches IPS minimum, 300 lb flange acess through 4 inches IPS alloy gate valve


GATE VALVE
MICROSWITCHES FOR TRANSVERSE CONTROL
INNER TUBULAR HOUSING
CONTROL VALVE AND OPERATOR
OUTER TUBULAR HOUSING
PROBE DRIVE HYDRAULIC CYLINDER
HYDRAULIC LINES
CONTROL UMBILICAL
CONTROL CONSOLE
HYDRAULIC SUPPLY SYSTEM
DOWTHERM COOLANT SYSTEMS AND CONTROLS
DOWTHERM COOLANT SUPPLY AND RETURN
SAMPLE LINE
TO VENT
DOWTHERM PUMP AND HEAT EXCHANGER

High temperature, high pressure sampling system.

Figure 9

ht134ee

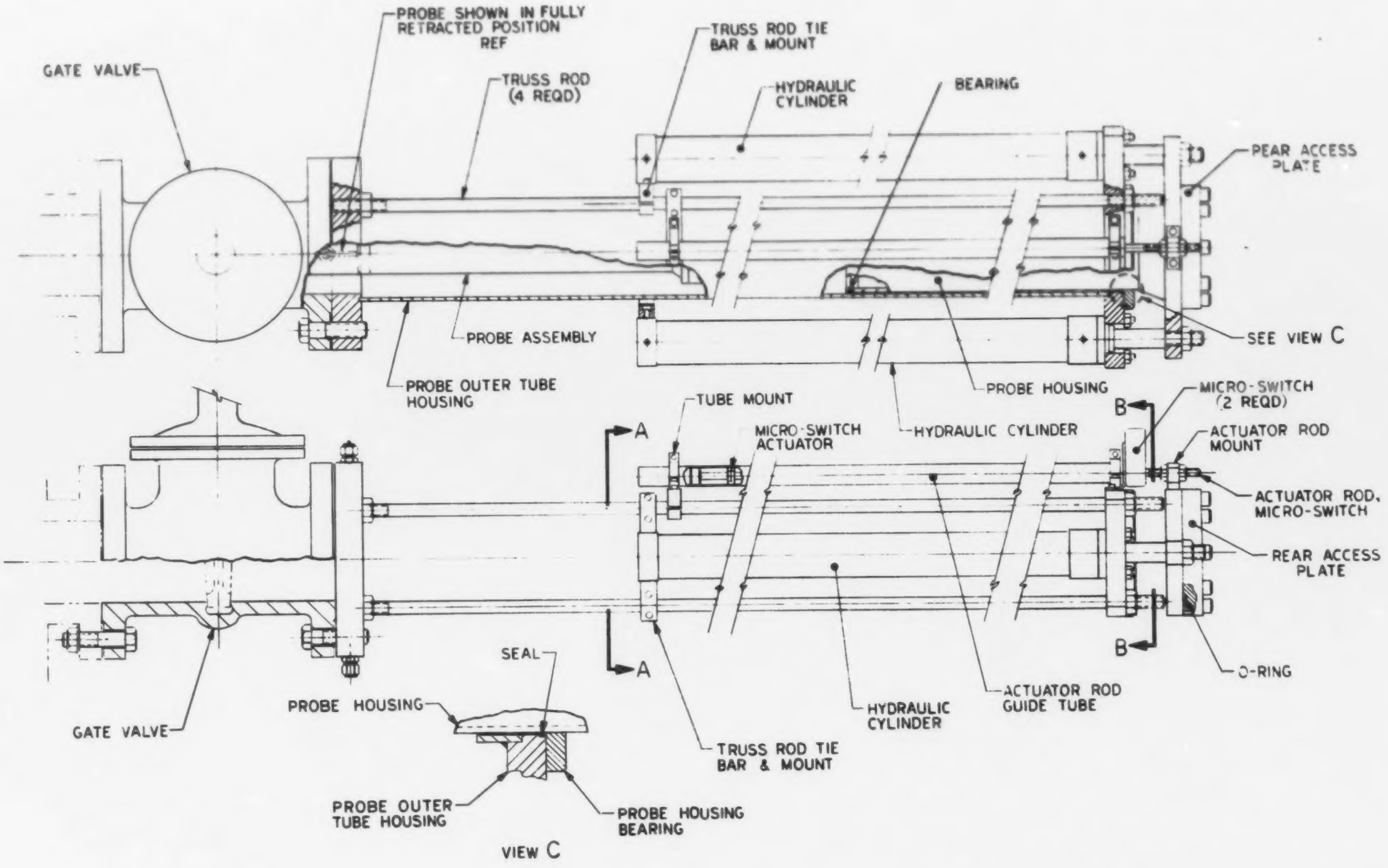
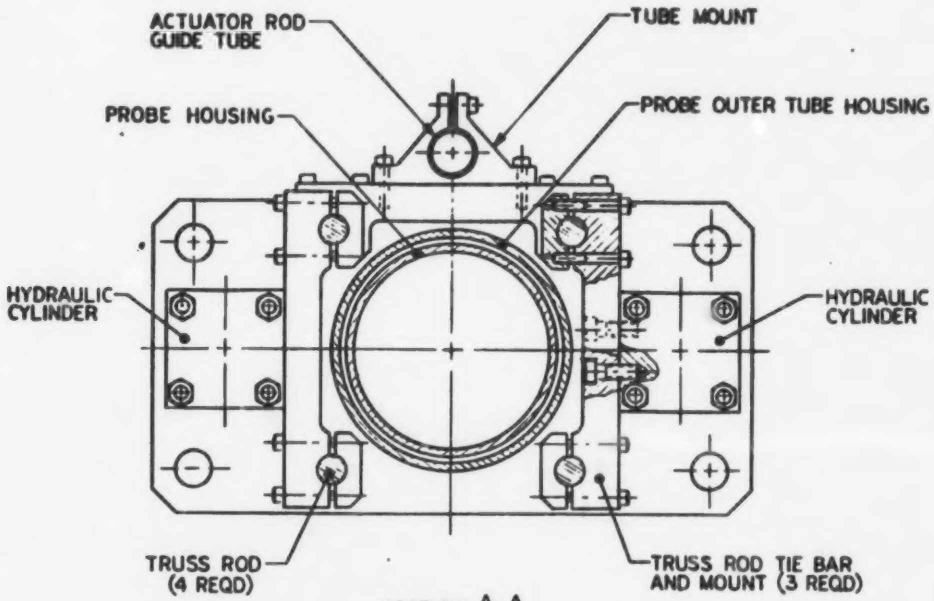
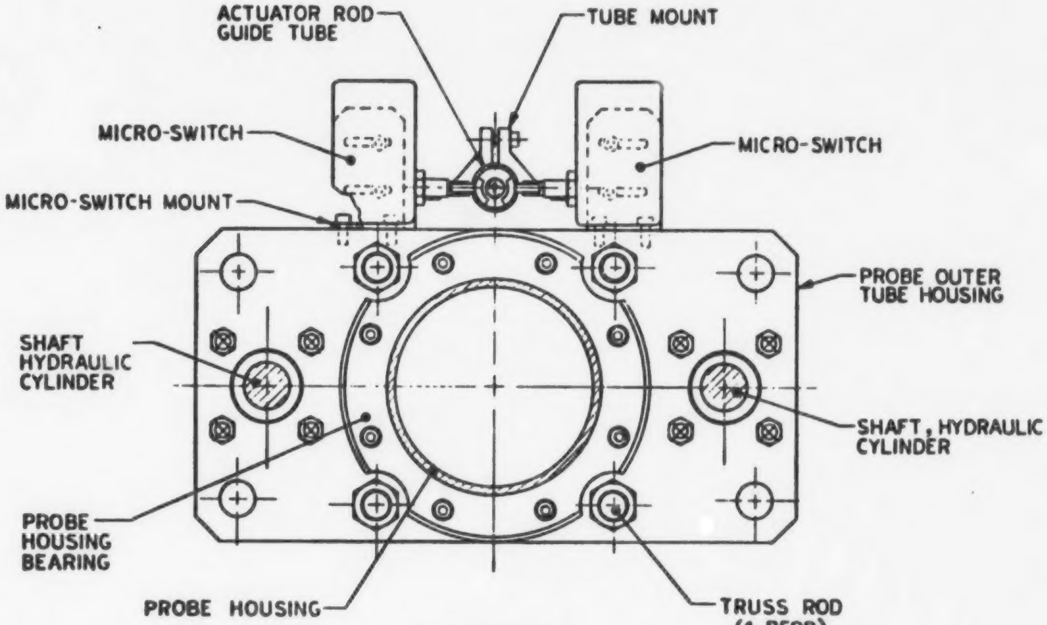


Figure 10(a) High temperature pressure probe assembly.



SECTION A-A



SECTION B-B

Figure 10(b) High temperature pressure probe assembly.

MEASUREMENTS FOR HIGH TEMPERATURE, HIGH PRESSURE PROCESSES

PROJECT ORGANIZATION

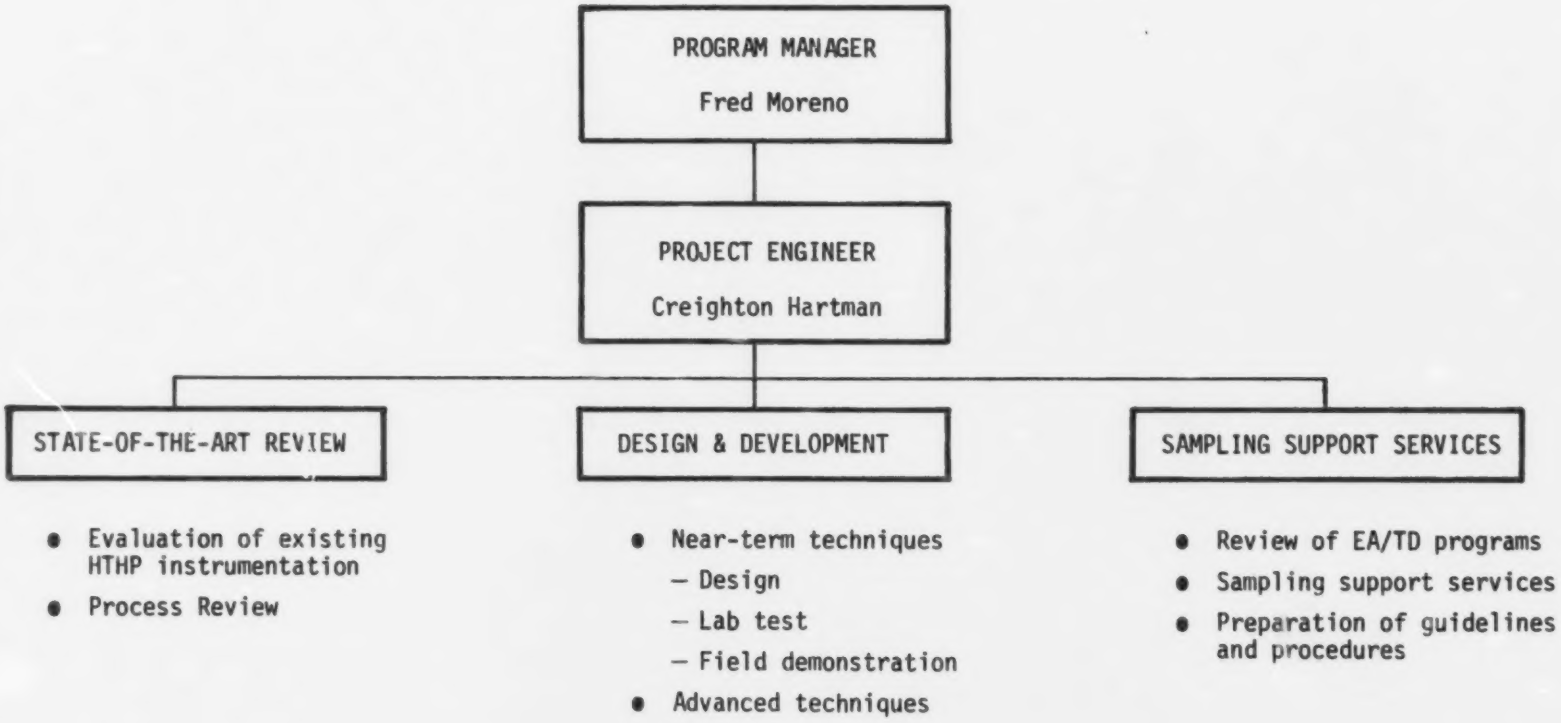


Figure 11 Program structure for EPA-sponsored HTHP process measurements program.

Pressure and Temperature

From a design point of view, the fundamental problems are those which arise as a result of the extremes of temperature and pressure. If the sampling system fails catastrophically when subjected to the duct environment due to inadequate structural integrity, other experimental problems (sticky particulate, probe deposition, cold wall condensation) or scientific niceties become totally irrelevent.

Two factors form keystone elements in any design related to HTHP sampling. These are:

- Safety — personnel operating the equipment should not be subjected to any substantial risk of injury to life or limb due to inadequate design
- Reliability — the device should operate over many cycles with no significant degradation of performance or increase of the probability of failure

The purely technical considerations for designing for extreme conditions have mostly to do with the structural behavior of materials. Critical material properties are severely degraded at temperatures above about 1000°F. At high pressures, this property degradation becomes critical because of the tremendous structural loads and stresses that can be imposed on the machine elements.

Seals are also critical. Insertion of a sampling device requires seals somewhere at the point of access. Safety and reliability both require there should be little possibility of a process leak. Even if thermally protected, a small leak of high temperature gas due to improper design, contamination of the sealing surface, or inadequate or careless seal installation, can lead to catastrophic consequences.

Attaining access to the high pressure environments for sampling is a challenging problem in itself. The larger the port required, the larger the blow-off loads on the system restraints. For a seal diameter of 4 inches, a representative port size for conventional stack sampling, the blow-off load is almost 20,000 pounds at 100 atmospheres!

To maximize safety and reliability in HTHP environments, sampling techniques, methodologies, and equipment should be well proven. The utilization of good pressure vessel design practices and thorough application and interpretation of analytical procedures cannot be overemphasized. Nondestructive tests, of which there are many types, should be utilized where possible to prove a sampling system or demonstrate basic integrity of critical component parts. Proof tests and simulation tests (where possible) in the laboratory create even greater confidence in a design.

In summary, safety and reliability can be maximized by:

- Thorough attention to detail
- Rigorous design and analysis
- Rigorous quality control of fabricated parts and certification of critical materials
- Thorough laboratory tests before field operations
- Detailed test planning, preparation, assembly procedures (leave nothing to chance)
- Thorough post-test inspections of equipment

Corrosion and Erosion

The environments associated with coal conversion technology can be extremely hostile. The most severe conditions exist, of course, in the main gasifier vessel and the transfer-lines which see high temperature, high pressure, corrosive gases, and high velocity char and ash particles which can cause severe erosion.

Fortunately, the sampling system designer generally does not have to deal with these problems on a long term basis as does the process designer. A material loss rate of 500 mils/1000 hrs may be acceptable to a sampling system used intermittently for short time exposures, whereas such a loss rate for a process component would be absolutely intolerable if reasonable lifetime is to be realized. However, some sampling components are sensitive to dimensional stability for proper operation (such as nozzle inlets, cyclones, inertial cascade impator plates); thus, materials performance becomes critical even for intermittent operation, and it is certainly true for sampling equipment used for in situ process monitoring, such as particulate monitoring upstream of a combined cycle turbine.

Both gaseous and liquid phase corrosion must be considered. Gaseous corrosion may consist of oxidation, carbonization, sulfadation and hydrogen embrittlement. Liquid corrosion may result from condensing liquid deposition from the gas onto surfaces below the dewpoint, by slag deposits from coal ash components, and by reactions of sulfur with nickel to form relatively low melting constituents in the metal.

There is currently little information available on erosion in high temperature, high pressure environments. Potentially, the most severe damage resulting from impact erosion is the removal of the protective oxide scale which protects from corrosion.

Space does not permit more in-depth review of these problems and even less so a discussion of their potential solutions. However, in general, it can be stated that a significant degree of protection from corrosive effects can be obtained by additions of nickel (>22 percent), chromium (>22 percent) and molybdenum (>3 percent) to the alloys. It has been found in some well planned experiments being performed by the Materials Properties Council of the American Gas Association that additional protection in gasification-type environments can be obtained by cladding with 671 alloy (50 percent nickel — 50 percent chromium) which is too brittle for most structural applications, by aluminizing, coating, and/or cooling.

Sometimes the erosion problem can be reduced by hard facing the metal with ceramic coatings. While this does not necessarily reduce corrosion, it will protect the oxide scale from erosion and thus the corrosion rate will not be accelerated by erosion.

Applications expertise in materials is critical to success. For example, many ceramics (which show great promise in corrosive/erosive environments) lack thermal shock resistance, which can lead to total failure by disintegration if not properly restrained or configured. Expense and lack of fabricability are factors which often mitigate against the use of exotic materials no matter how great their potential. Ultimately, the best design advice is (where possible):

- Cool components to reduce operating temperature
- Reduce flow velocities, especially particle-laden streams
- Design for low stresses

Condensation of Process Constituents

Many HTHP processes under consideration for particulate and gas sampling contain many chemical species which may be liquids or solids at room temperature. Therefore, any proposed sampling approach must consider the possibility that under certain conditions some species may change physical states, even several times, during the sampling process.

For example, many of the "coal tar" derivations (cresol, phenol, toluene) have critical pressures in the region of 40 to 50 atmospheres. Their critical temperatures are above 300°C. If a sampling probe operating at 500°C and 70 atmospheres in a coal processing stream (gasification, liquification, etc.) samples an appreciable volume of those components and cools them below 300°C (which would be reasonable) before reducing the pressure, these species are quite likely to condense and plug the flow channels with spurts of liquids. Even worse, if the pressure is then reduced after cooling, these components can later flash from liquid to gas before finally being collected.

Condensation can be a significant problem in the "simple" combustion environment. Streams with very high percentage levels by volume (20 percent) of water vapor can have condensation temperatures as high as 300°F at pressures of 300 psia. However, gasification probably represents the most severe potential problems of high temperature condensation. A LURGI gasifier, for example, produces a broad variety of tars (boiling point 500°F+), tar oils (B.P. 285°F - 680°F), and naptha oils (B.P. 120°F - 320°F). If sample integrity is to be preserved and a representative sample obtained in such systems, the sample must be held at very high temperatures indeed; quite possibly as high as 750°F to 1000°F in some cases.

Compounding this problem are those constituents such as phenol, resourcinal, napthalene, etc., which have relatively high vapor pressures at low temperatures. It would be very easy to "lose" these compounds during sample handling.

To illustrate this problem more explicitly, Figure 12 shows how high temperature condensibles could affect the sampling process. The process stream conditions are taken from the Clean Fuels from Coal Symposium, September 1973, sponsored by IGT, Chicago. For illustration, worse case conditions were assumed to prevail, and a poorly designed, but not totally unrepresentative, sampling system was assumed.

Clearly, the chemical composition of the process stream must be considered in the selection of sampler operating conditions, not only in the sampling probe itself, and in the particulate removal stages, but in all the subsequent gas sample steps which follow.

Toxicity of Process Constituents

Sampling personnel should be aware of the potential health hazards that exist with respect to many of the process constituents in HTHP processes. Many of the "coal tar" compounds found in gasification processes are known carcinogens or have other toxic properties, which makes operating, cleaning, and maintaining the sampling equipment a safety consideration.

Flammability and Explosion Hazards

In gasification processes, the gas flow through the sampler will be flammable and/or explosive. Significantly, sampling rates are such that after expansion, considerable volume is represented which may preclude simple venting if there is any possibility of sparks, static charge build-up, or open flame.

Certainly, sampling equipment for such processes should be grounded, and all motors, switches, and controls should be explosion-proofed. Whatever alternatives are selected for handling these gases, protection of personnel and process must be of primary consideration.

CONCLUSION

It is anticipated that over the next several years important developments will be forthcoming in coal conversion technology, as the effort continues to utilize this nation's abundant coal reserves in an environmentally acceptable manner. A necessary component of this development will be the

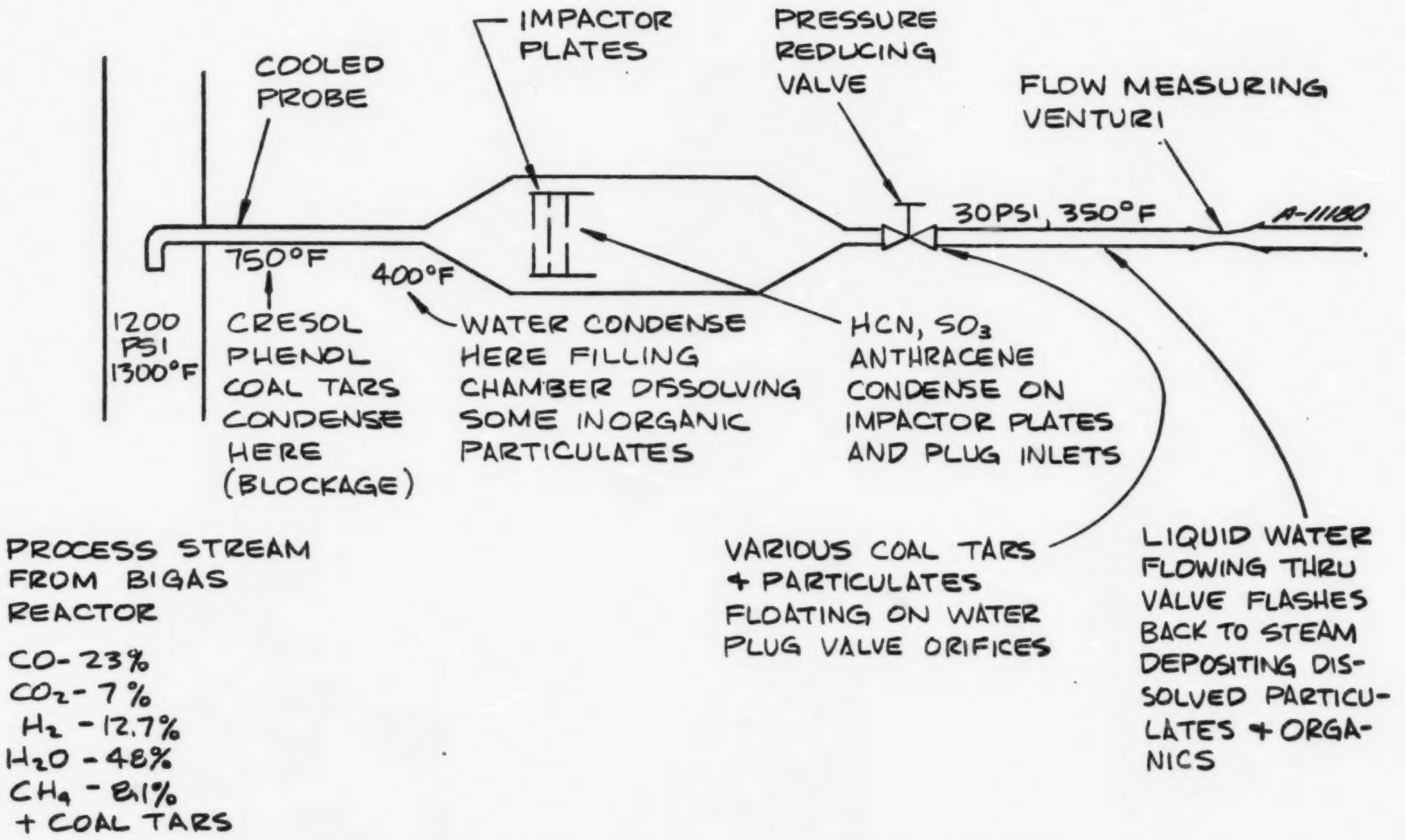

IMPACTOR PLATES
PRESSURE REDUCING VALVE
COOLED PROBE
FLOW MEASURING VENTURI
30PSI, 350°F
A-11180
750°F
400°F
1200 PSI 1300°F
CRESOL PHENOL COAL TARS CONDENSE HERE (BLOCKAGE)
WATER CONDENSE HERE FILLING CHAMBER DISSOLVING SOME INORGANIC PARTICULATES
HCN, SO3 ANTHRACENE CONDENSE ON IMPACTOR PLATES AND PLUG INLETS
PROCESS STREAM FROM BIGAS REACTOR
CO- 23%
CO2- 7%
H2 - 12.7%
H2O - 48%
CH4 - 8.1%
+ COAL TARS
VARIOUS COAL TARS + PARTICULATES FLOATING ON WATER PLUG VALVE ORIFICES
LIQUID WATER FLOWING THRU VALVE FLASHES BACK TO STEAM DEPOSITING DISSOLVED PARTICULATES + ORGANICS

Figure 12 Condensation problems arising from hypothetical inadequate sampling approach.

evolution of new particulate and gaseous sampling instrumentation, methodologies, and techniques. This paper has presented only briefly some of the challenges that face this evolutionary development.

FBC PARTICULATE CONTROL PRACTICE AND FUTURE NEEDS: EXXON MINIPLANT

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EXXON Research and Engineering Company
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FBC PARTICULATE CONTROL PRACTICE AND FUTURE NEEDS: EXXON MINIPLANT

R. C. Hoke

Exxon Research and Engineering Company
Linden, NJ

ABSTRACT

The successful development of the pressurized fluidized bed coal combustion system is dependent on the removal of particulates from the high temperature, high pressure flue gas before it is allowed to expand through a gas turbine. Current estimates of the allowable particulate loading at the turbine inlet, based on turbine blade erosion limits, are in the range of 0.001 to 0.05 gm/m^3 (0.0004 to 0.02 gr/SCF). These limits are lower than the current EPA particulate emission standard of 0.115 gm/m^3 (0.05 gr/SCF). Corrosion of the turbine by alkali sulfates is also a potential problem and may impose even tighter allowable inlet particulate loadings.

Reducing particulate loading down to these low levels requires very efficient devices capable of operating at higher temperature (950°C) and high pressure (10 atm). A number of devices were considered and the system chosen consists in two stages of cyclones followed by a high efficiency granular bed filter. The granular bed filter must be capable of 90 to 99% removal efficiency to meet the estimated particulate loading requirements. Although granular bed filters are still in the development stage and have not as yet demonstrated these high removal efficiencies, they offer the most promise compared to other available devices. A granular bed filter has been purchased from the Ducon Company to be installed on the Exxon FBC miniplant. A test program will be carried out to determine particulate removal efficiency, maintenance of removal efficiency and mechanical performance.

FBC PARTICULATE CONTROL PRACTICE AND FUTURE NEEDS: EXXON MINIPLANT

Fluidized bed combustion (FBC) is a new direct concept technique which permits the use of coal in an efficient manner with minimum impact on the environment. Two versions are being developed. The first operates at atmospheric pressure and is being developed for use in both electrical power generation and production of steam for industrial use. The second operates at higher pressure, up to 10 atm absolute, and is being developed for use in a combined cycle power generation system. A sketch of a pressurized FBC system is shown in Figure 1. Coal, along with limestone or dolomite which act as SO_2 sorbents, is injected into the bottom of the pressurized boiler. Coal is burned in the limestone bed which is fluidized by the incoming combustion air. SO_2 formed in the combustion process is removed by the limestone bed. Steam coils immersed in the fluidized bed remove the heat of combustion and maintain the bed temperature in the range of 1500 to 1700°F. Steam thus generated, is used to operate a steam turbine. The desulfurized flue gas passes through a particulate removal system and is then expanded across a gas turbine which generates additional power and also operates the combustion air compressor. A portion of the particulates removed from the flue gas is recycled to the boiler to increase the carbon combustion efficiency. The particulate removal system must reduce the particulate loading down to levels sufficiently low to protect the gas turbine and meet current emission standards.

Figure 1 also shows a regeneration section coupled to the boiler. This is an adjunct to the boiler which regenerates the sulfated limestone to lime which is then returned to the boiler. An off gas with SO_2 concentration sufficiently high to be fed to a sulfur recovery system is also produced. The regeneration system is currently at an earlier stage of development than the boiler system. Although an FBC system can operate with limestone used in a once-through manner, the regeneration of sulfated limestone would reduce the quantity of limestone fed to the boiler and the disposal of used limestone.

An atmospheric pressure FBC system is similar to the pressurized system with the exception of the flue gas turbine which is not used.

The FBC system offers a number of advantages over a conventional coal fired boiler. It effectively controls SO_2 emissions without external scrubbing systems. NO_x emissions are also very low because of low combustion temperatures and also because conditions in the boiler favor NO destruction reactions. For example, under pressurized FBC conditions, NO_x emissions are typically in the range of 100 to 150 ppm. An FBC system can also handle a variety of fuels, gaseous, liquid and solid. It also provides higher generation efficiency at lower capital and operating costs.

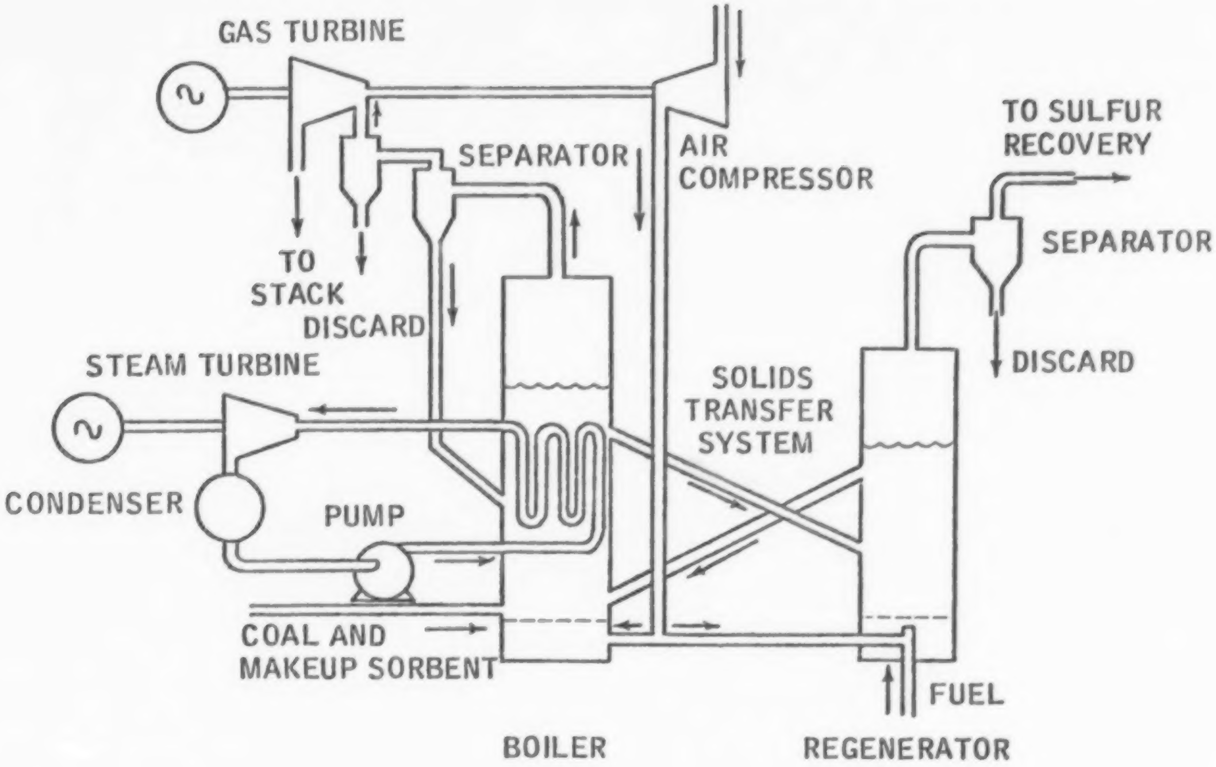


Figure 1

PRESSURIZED FLUIDIZED BED COAL COMBUSTION SYSTEM

The work currently in progress at Exxon Research and Engineering Company is sponsored by the U.S. Environmental Protection Agency (EPA). A number of other FBC programs are now underway in the U.S. and the U.K. under sponsorship of the EPA, the U.S. Energy Research and Development Administration (ERDA) and the Electric Power Research Institute (EPRI). These programs have progressed from the laboratory and bench scale to process development units burning 5 to 10 T/D of coal. Larger pilot plants, of atmospheric and pressurized design in the range of 10-30 MWe, are now under design and construction. Environmental assessment and economic evaluation studies are also in progress.

The program at Exxon Research is being carried out in two pressurized FBC units, a bench scale unit and a process development unit. The process development unit, also known as the "miniplant" is capable of burning up to 6 T/D of coal at pressures up to 10 atm. The design of the unit was described in a report submitted to the EPA[1]. The combustor section of the miniplant has been in operation since July 1975 and has to this date logged over 1200 hrs of coal combustion time. This includes a 10 day continuous demonstration run which was completed in November 1975.

One of the objectives of the FBC program at Exxon Research is the evaluation of particulate removal systems. These systems must satisfy three performance criteria as shown in Table 1.

Table 1

Particulate Emission Control Requirements

Requirement	Allowable Level (gr/SCF)	Required Efficiency of Third Stage Device (%)
Environmental	0.05	67
Turbine Erosion		
Range of Estimates	0.02-0.0004	87-99.7
Tentative Level	0.002	98.7
Turbine Corrosion	?	?

The first performance requirement is based on the environmental emission standard as set by the EPA. The second is protection of the gas turbine from excessive erosion and the third is protection of the turbine from particulate induced corrosion. The environmental emission standard is 0.1 lb particulate/10^6 BTU fired. This is equivalent to a particulate concentration in the flue gas of about 0.05 gr/SCF. At the present time, turbine erosion limits are not well defined. Westinghouse Research Laboratory recently estimated allowable levels based on limited data and model studies.[2] The estimates covered a range of 0.02 to 0.0004 gr/SCF. Based on these estimates, Westinghouse suggested a tentative allowable level of 0.002 gr/SCF. Insufficient data are available to estimate the allowable level to prevent corrosion, but some qualitative estimates suggest that these levels may be even lower than the erosion limits.[3] The particulate concentrations determined by the environmental and erosion requirements are lower than can be met with simple particulate removal systems such as high efficiency cyclones. It is estimated that a two stage cyclone system could reduce the particulate loading to a level of about 0.15 gr/SCF. Therefore, a third stage particulate removal device would be also required to reach the levels set by the environmental and erosion requirement. The efficiency of such a third stage device is also shown in Table 1. To meet the environmental requirements, an efficiency of 67% would be required. Efficiencies of 87 to 99.7% would be required to meet the range of particulate levels set by erosion limits. If the tentative limit suggested by Westinghouse is used, the required removal efficiency would be 98.7%. Obviously, a very efficient third stage system will be required.

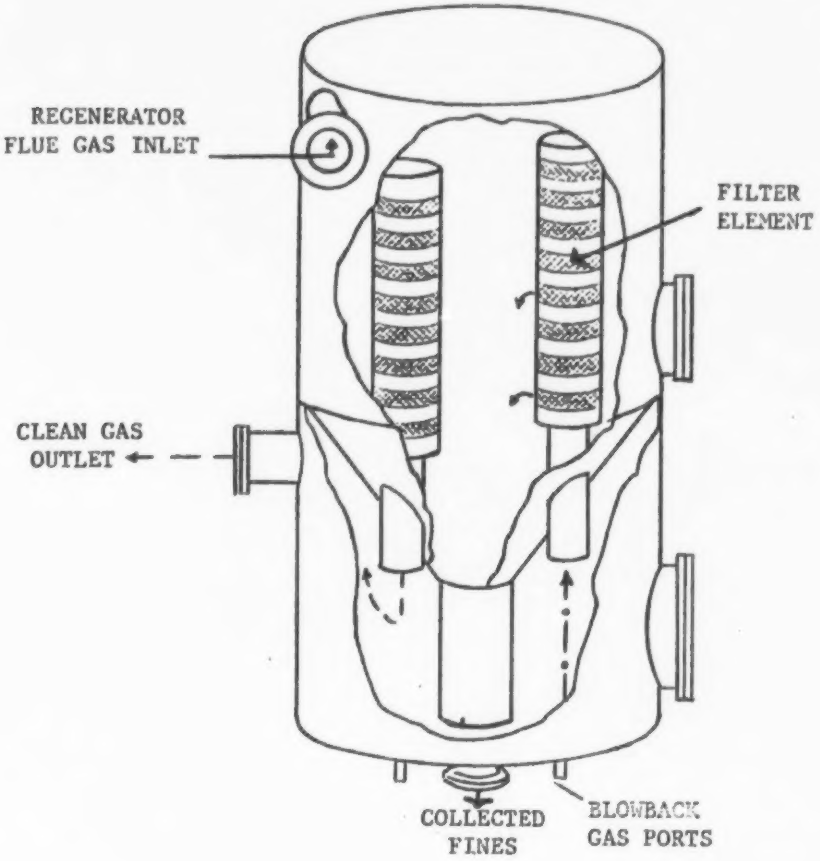
Some surveys have been made of the type of particulate removal system which may be capable of satisfying these requirements. Stone and Webster[4] and Westinghouse[2] have published such surveys. Other evaluation programs sponsored by EPA, EPRI and ERDA, are also now underway. Three systems have been mentioned in these studies which could provide the required degree of particulate removal. These are granular bed filters, high temperature metal or ceramic filters and a low temperature scrubber combined with an efficient heat exchanger. The consensus appears to favor granular bed filters as the type which offers the best chance of meeting the required high removal efficiencies.

A number of groups are currently developing granular bed filter systems and are described in the above referenced Westinghouse and Stone and Webster reports. The Ducon Company has tested granular bed filters on refinery and other waste gas streams. A type of filter, called the panel bed filter is currently being studied at the City College of New York under sponsorship of EPRI. The Rexnord Company has installed a number of "gravel bed" filters on cement kiln and other industrial off gases. Combustion Power Company has also installed a number of "dry scrubber" filters on flue gas from wood waste boilers. Combustion Power is also developing a similar system for application on FBC.

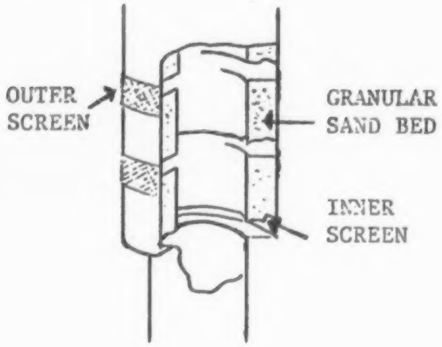
A sketch of the Ducon filter is shown in Figure 2. The filter consists of a series of beds containing the filter medium, stacked vertically to form a filter element. A number of elements are contained in a pressure shell. Dirty gas passes through an inlet screen, through the filter medium and out through an outlet screen into a central collecting tube. Clean gas exits the vessel at the bottom. Each element is periodically cleaned by a short pulse of high pressure clean air flowing in reverse flow through each element. The dust is blown out through the inlet screens and collects in the bottom cone of the pressure vessel. The filter medium is retained in the beds.

Figure 2

DUCON GRANULAR BED FILTER



FILTER ELEMENT INTERNALS



The CCNY panel bed filter is described in Figure 3. Thin sections of the filter medium are formed between louvred panels as shown in cross section in Figure 3. Four panels would be placed at right angles to form a hollow square cross section. Dirty gas is filtered by passing through the granular filter medium. The medium is periodically cleaned by a "puff-back" of clean air in the reverse direction. This blows off the dust and some of the filter medium which is collected, cleaned externally and recirculated to the top of the bed. After each puff-back, cleaned medium drops into the panel bed to replace material removed by the puff-back.

A sketch of the Rexnord gravel bed filter is shown in Figure 4. Dirty gas enters into a cyclone pre-cleaner which is an integral part of the system. The pre-cleaned gas passes out the vortex tube and passes down through the gravel beds and out the clean gas exit. The beds are periodically cleaned by a reverse flow of clean gas. During the reverse flow step, the gravel beds are raked by a rotating raking mechanism to aid in removing the dust from the filter medium. The dust is then blown in reverse direction through the cyclone and collects at the bottom of the pre-cleaner.

Figure 5 is a sketch of the Combustion Power dry scrubber. Dirty gas enters the containing vessel and flows through a louvred anular vessel containing the filter medium. Clean gas exits through the central collector. The filter medium is constantly fed into the top of the vessel, continually moves down through the louvred containing vessel, passes out the bottom, is cleaned externally and recycled back to the inlet.

The Ducon filter was chosen for testing on the Exxon/EPA miniplant after considering all the available systems. The selection was based on previous experience with the Ducon system which indicated that it had the potential of providing high removal efficiency. Another desirable feature of the Ducon system is the retention of the granular filter medium in the filter vessel. In all other systems, the medium is removed, cleaned externally and recycled back to the filter vessel. The Ducon system to be used on the miniplant will consist of four filter elements, each 1 ft in diameter by 6 ft long. The elements will be enclosed in a refractory lined pressure vessel designed for 10 atm operating pressure. The system will be capable of filtering 1200 SCFM of hot flue gas. It is currently being fabricated and will be installed late in 1976. It will be evaluated in an experimental program in which the particulate removal efficiency and efficiency maintenance will be measured as a function of operating parameters. Equipment life will also be studied. A cooperative program is also being developed by EPA, ERDA and a gas turbine manufacturer to install and test samples of candidate gas turbine materials in the filtered flue gas. A second particulate removal system will also be selected, installed and tested as part of the EPA program after the evaluation of the Ducon filter is completed.

Figure 3

CCNY PANEL BED FILTER

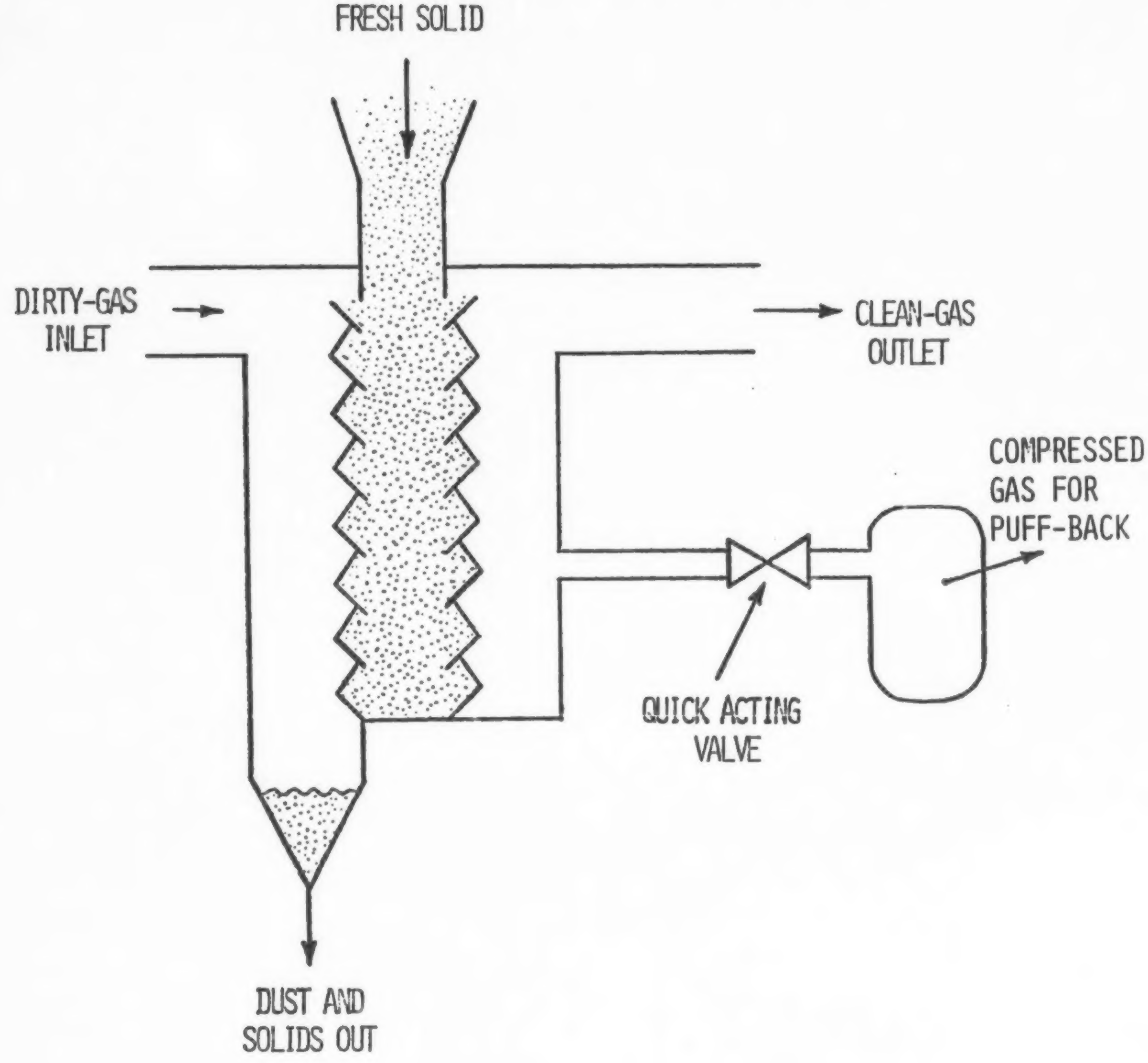


Figure 4

REXNORD GRAVEL BED FILTER

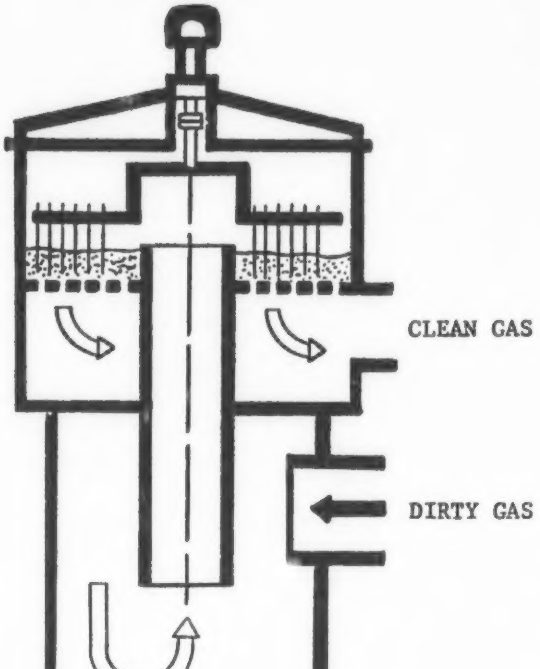
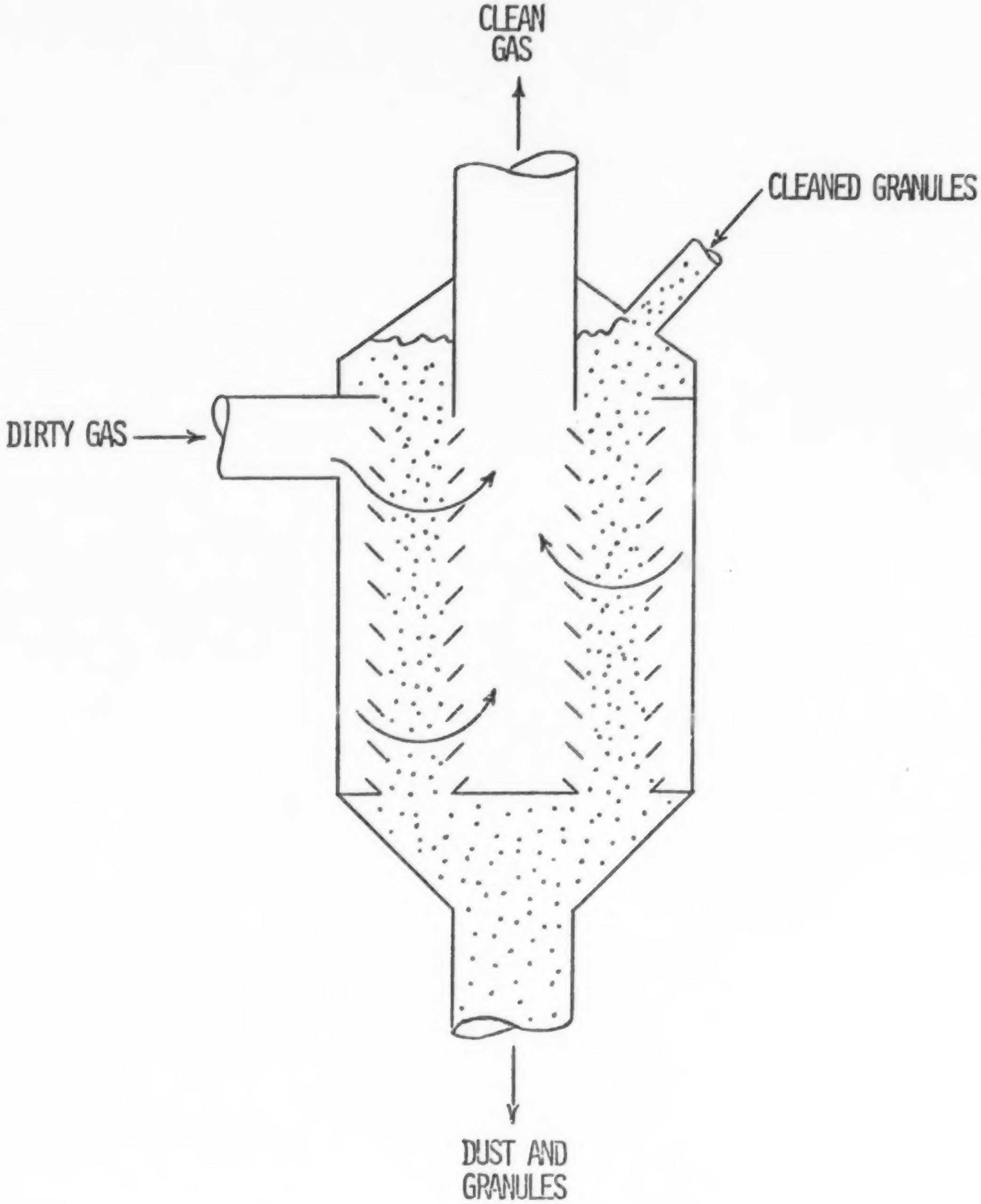


Figure 5

COMBUSTION POWER GRANULAR BED FILTER



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PARTICULATE CONTROL REQUIREMENTS FOR GASIFIED COAL-COMBINED-CYCLE POWER PLANTS

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ABSTRACT

The combination of low-Btu gasification processes with combined-cycle power generation offers an attractive means of generating electricity from high-sulfur coal while minimizing pollutants. Because the mass flow rate of low-Btu gas is a significant fraction of the combustor air flow, the ability to use the large amounts of sensible heat contained in the fuel gas at combined-cycle efficiency is a key factor in system performance. Best performance would be obtained by using the gas without cool down after leaving the gasifier. However, this means that all cleanup systems, including those for particulates must operate at extremely high temperature while producing a gas of a quality that is acceptable to the gas turbine and that will meet effluent requirements. In order to identify the particulate removal problem it is necessary to define the particulate content of the gas entering the cleanup system, as well as limits imposed by the turbine and the environment.

To realistically assess a candidate particulate removal system it is necessary to consider the alternatives. The conventional approach to the problem would be to reduce gas temperature to a level compatible with a water scrub. The low-temperature gas could then be used directly in the burner or reheated by regeneration against the dirty gas to minimize the performance decrement. The performance and cost associated with these alternatives are presented and their interaction with other cleanup functions, particularly those for sulfur compounds and ammonia, are discussed in an attempt to define the benefits that can be expected to offset the cost of high-temperature particulate removal.

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PARTICULATE CONTROL REQUIREMENTS FOR GASIFIED COAL-COMBINED-CYCLE POWER PLANT

INTRODUCTION

Our ever increasing reliance on foreign sources for energy can be relieved to some extent by changing the manner in which electricity is generated. At present, about 40% of the electric power in this country is based upon the use of oil and natural gas, about 53% on coal and 7% on nuclear energy. While the U.S. has vast amounts of coal reserve, much of the coal near the major urban areas contains too much sulfur to be burned directly. Of the several ways of generating nonpolluting power using coal currently under investigation, gasification in conjunction with an advanced power system such as the COmbined Gas and Steam (COGAS) system appears to be one of the more attractive possibilities.

The COGAS system (Fig. 1) is based upon the use of advanced, high-temperature gas turbines such as those being pursued under a recently announced ERDA program to develop the technology for 2600 F turbine inlet temperature machines. Like present day utility gas turbines which operate at under 2000 F, these advanced turbines will require quite clean fuels; fuels which have occasionally been referred to as being of pharmacutical grade.

This cleanliness is required for several reasons, the major ones being erosion and corrosion. Many of the turbine blades have very thin coatings of oxidation resistant materials on their surface. Thus, not only the problem of catastrophic erosion which changes the blades aerodynamics must be considered, but erosion by small particles which could abrade the coatings would now be a problem. Once the coatings have been penetrated, alkali metal salts, usually sodium or potassium sulfates, attack the high-strength, high-temperature alloys which form the blade.

To keep blade metal temperatures to reasonable levels, below 1600 F, some type of cooling is used (Fig. 2). One method uses small diameter holes (0.03 to 0.05 in.) in various parts of the blade to release cooler air from the blade interior to form a film of cooling around the blade. It is possible that deposition of solids could block these holes and cause hot spots thereby leading to localized blade failure.

Thus, it is important that substances which are not only harmful to the environment, but also harmful to the power system be removed or reduced to acceptable levels. It still remains, however, to fully define the scope of the problem.

CLEANUP SYSTEM GOALS

The off gas from a low-Btu coal gasifier contains CO, H_2, N_2, CO_2, and H_2O as its major constituents. Unfortunately, the sulfur content in the coal is

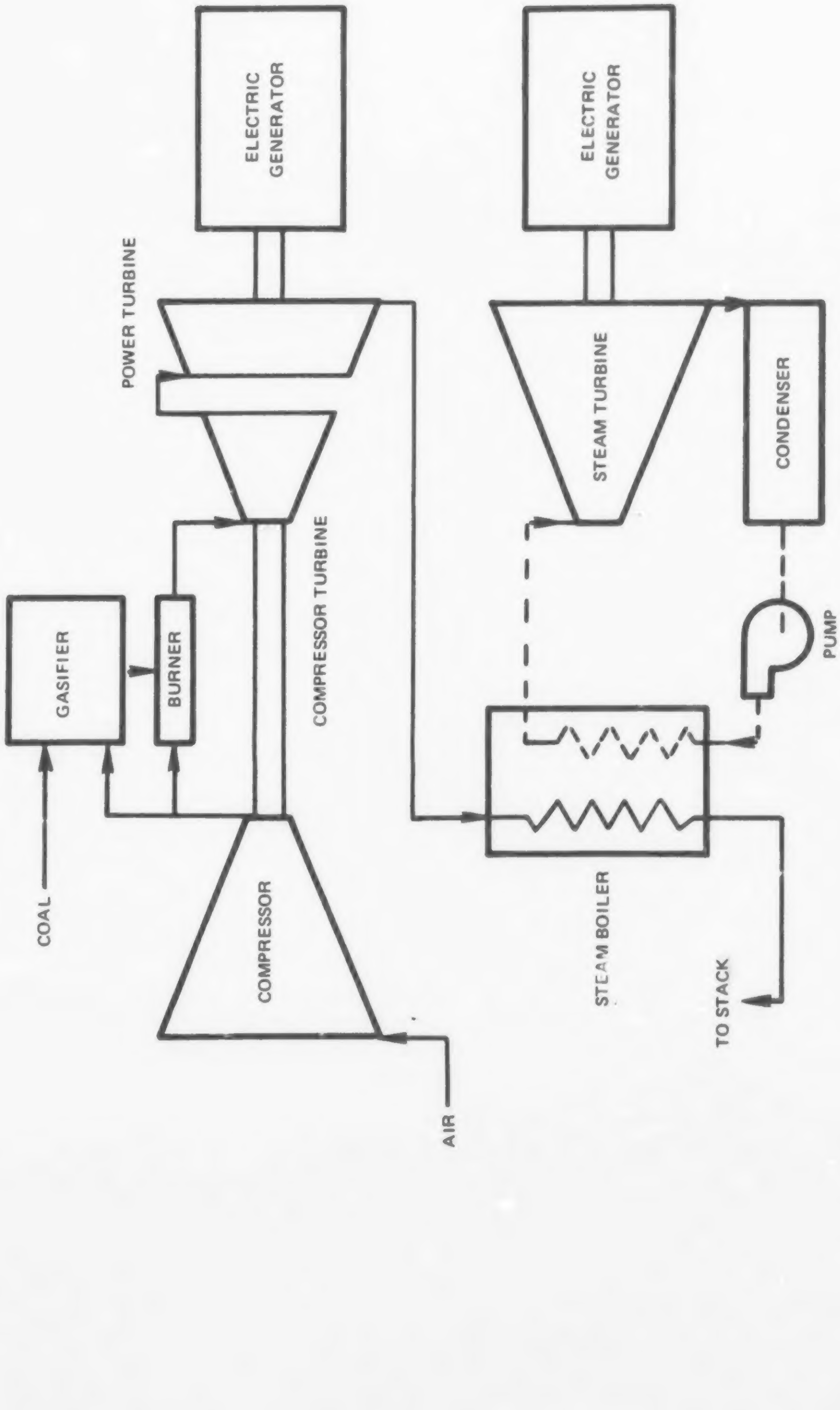

Figure 1: Combined Gas–Steam Turbine System

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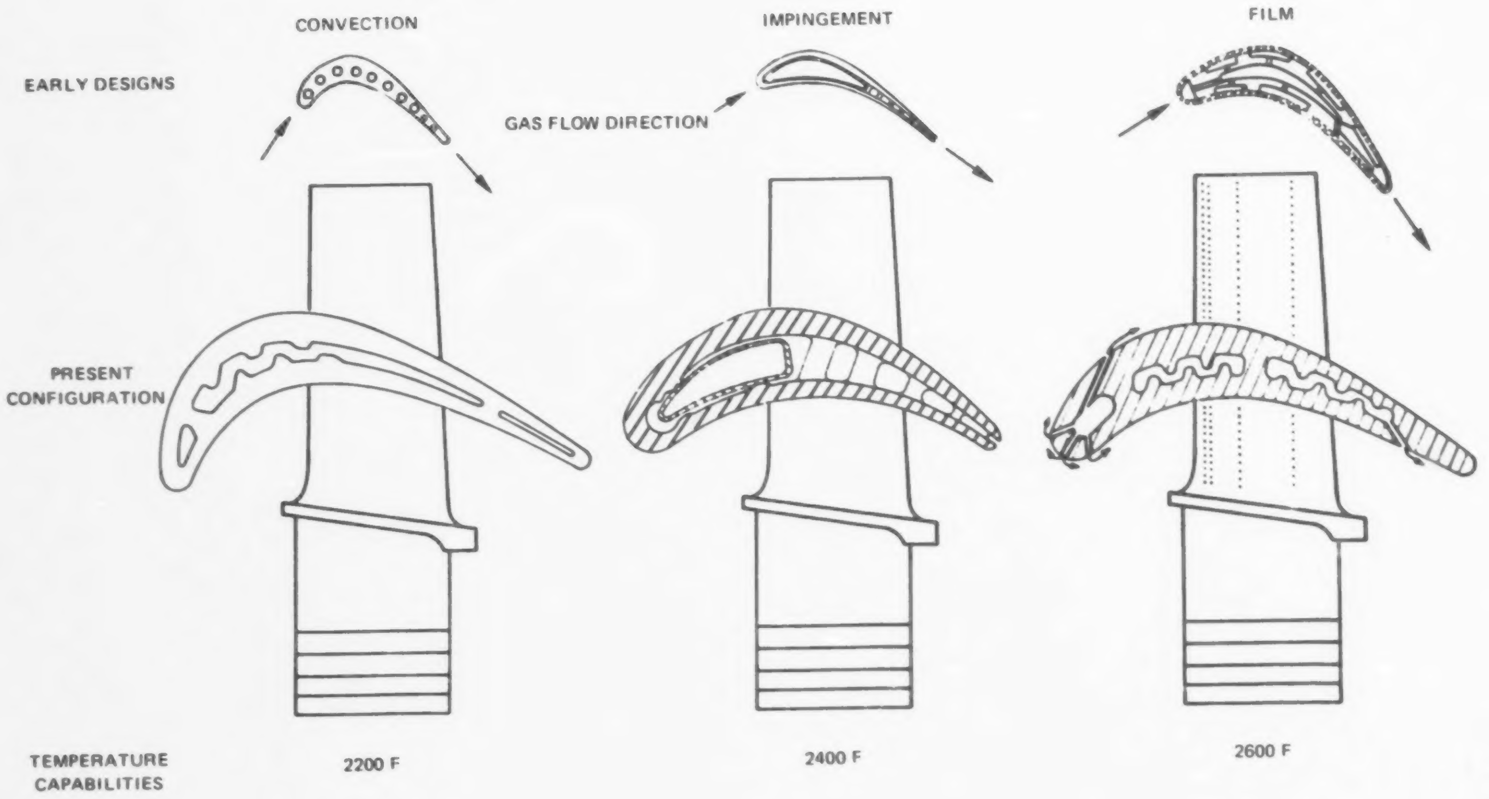

Figure 2: Turbine Blade Cooling Configuration

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onverted mainly to H_2S and other sulfur bearing compounds, and in some types of gasifiers, the organically bound nitrogen in the coal is converted to ammonia which, upon combustion would form NO_x. Of course, coal ash, carbon carryover, and other solids will be carried in the off gas. Depending upon the gasifier type, this gas may be between 1000 F and 3000 F and between 1 atm and 40 atm. For example, the system to be described in the following sections has a two-stage gasifier with the off gas exiting at 1800 F and approximately 30 atm. Using Illinois No. 6 coal, the off gas contains 0.51 mol percent sulfur compounds, 0.38 mol percent nitrogen compounds and approximately 0.8 gr/SCF of particulate. (A solids carryover equivalent to 10% of the ash content has been assumed. Actual values of particulates from various gasifiers have yet to be verified.)

The size distriubtion of the particulates from the gasifier has not been adequately documented. Since pulverized coal is the fuel to the gasifier, it has been assumed that the size distribution in the ash carryover would resemble that of the fly ash from a P.F. boiler as shown in Table 1.

The goals for cleanup given in Table 2(1) are based upon both environmental and turbine requirements. The present EPA standards for large coal-fired stations of 1.2 lb SO_2/per 10^6 Btu, 0.7 lb NO_x/per 10^6 Btu and 0.1 lb particulate/per 10^6 Btu are easily met if these goals can be attained.

CLEANUP SYSTEM IDENTIFICATION

The gas from the gasifier can be cleaned to the levels indicated in the suggested goals using commercially available processes. However, these processes operate at low temperatures (below 250 F) and thus require considerable fuel gas cooling. The various irreversibilities during this cooling process impose a thermodynamic penalty. Also, higher process utility loads for the low-temperature process results in an energy penalty. The attractiveness then of the high temperature cleanup system arises from the potentially better performance that the overall integrated system (gasifier/cleanup/COGAS) might have. To demonstrate this, the performance and cost of the low- and high-temperature system will be compared.

Integrated Power System with Low-Temperature Cleanup Process

An integrated power system using a low-temperature cleanup process is shown schematically in Fig. 3. The cleanup process is the Allied Chemical Corporation Selexol system and is capable of reducing the sulfur compounds to the 100 ppm level; and with additional water wash, reducing the ammonia content to acceptable levels. Since this process involves aequeous scrubbing in a number of stages, the particulates are removed to very low levels at a cost and performance penalty that is not easily differentiated from the sulfur/ammonia removal process.

A brief description of the salient operating characteristics of this system are given in Table 3. The cost and performance are given in Table 4 and Table 5, respectively.

TABLE 1

ESTIMATED PARTICLE SIZE DISTRIBUTION

(Flyash From P. F. Boiler)

Range	Comulative Percent (wt)
75µ+	11
50-75	19
25-50	54
10-25	82
5-10	94
2-5	98
< 2	100

TABLE 2

FUEL GAS CLEANUP

	Low-Btu Gas	Typical Current Spec
Sulfur	0.05 mol% or less than amount to form 0.6 ppm alkali metal sulfate	< 1.0 mol% or less than amount to form 5 ppm alkali metal sulfate
Particulates	4 ppm wt or 0.0012 gr/ft^3 > 2μ	30 ppm or 0.01 gr/ft^3
Metals		
Vanadium	< 0.003 ppm wt	< 0.002 ppm wt
Sodium/Potassium	See sulfur spec	< 0.06 ppm
Nitrogen	500 ppm as NH_3	---

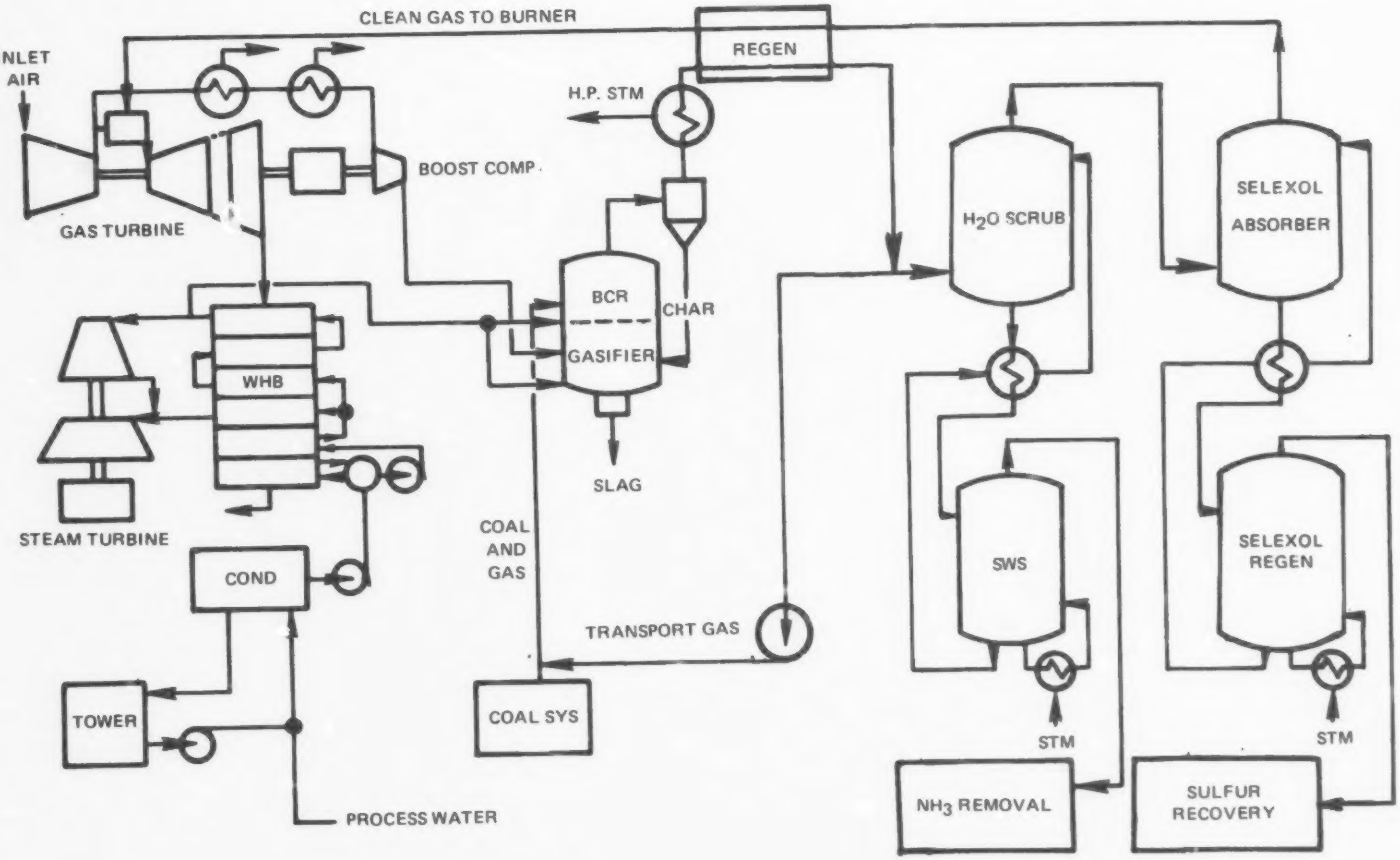

Figure 3: BCR/Selexol System

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TABLE 3

SYSTEM CHARACTERISTICS

Power System	Gas Turbine	Steam
· Temperature - F	2600	1000
· Pressure - psia	350	1250
· Unit Size - MW	180 (Nom)	300 (Nom)
· Gasifier		
· Exit Temperature - F	1800	
· Pressure - psia	400	
· Fuel Gas HHV-Btu/SCF	135-150	

TABLE 4

SYSTEM PERFORMANCE

	Low-Temperature	High-Temperature	High-Temperature/ww
Gas Turbine Out-MW	734	858	770
Steam Out-MW	321	297	293
Auxiliaries-MW	82	77	78
Net-MW	973	1078	985
Efficiency	38.6%	42.7%	39.0%
Emissions lb/10^6 Btu			
SO_2	0.39	0.55	0.55
NO_x	0.29	5.5	0.8
Particulates	0.01	0.01	0.01

TABLE 5

SYSTEM COST

(Millions of Dollars)

	Low-Temperature	High-Temperature	High-Temperature/ww
Gasifier	92	92	92
Desulfurization	27	23	23
Particulate Removal	---	27	---
Other Process Cost	82	44	82
Power System	222	247	227
Cost	423	433	424
Cost - $/kW	434	402	430
Power Cost-mills/kWhr	21.5	19.5	21.2

Integrated System With High Temperature Cleanup Process

Fig. 4 shows a schematic of an integrated system with a high-temperature cleanup process. The sulfur content is reduced to about 500 ppm in a CONOCO (CONOCO Coal Development Center) half-calcined dolomite fluid-bed reactor which operates at 1600 F. This system is one of several types of sulfur cleanup processes which operate at high temperature that are currently in the pilot stage. In this process, as in most high-temperature processes there is no mechanism for removal of fuel-bound nitrogen compounds.

The high-temperature particulate removal system, shown schematically in Fig. 5, consists of three separate units. The first unit would be a high efficiency cyclone of a more or less conventional design, but constructed of high-temperature alloys. Approximately 85% of the particles above 20μ could be removed in this device. Following the cyclone, a series of small diameter cyclones or multiclones, also of high-temperature alloys would remove 90% of the remaining particles above 10μ. The 'final filter' would be selected from the various granular bed, wire or ceramic mesh, or other devices briefly described in Table 5. These devices would remove 99% of the residual particulate above 2μ.

The actual performance of such devices has not been well established at the temperatures and pressure of interest. Work at the Argonne National Laboratories(2) on pressurized fluid beds operating at 10 atm and 1600 F has shown that cyclones followed by two stages of metal mesh achieved 99%+ removal of micron sized particles from a slipstream. Similarly, work at CONOCO Coal Development Center(3) indicates that metallic mesh removed not only particulates, but that it could potentially remove alkali metal vapor aerosols. It must be cautioned that the foregoing results are based upon small-scale tests. It remains to be seen if these devices can be scaled to the required sizes (~ 10^5 ACFM).

It is difficult to project costs for systems which have yet to leave the pilot stage, but it would appear that high temperature particulate cleanup is not an inexpensive problem. It was mentioned before that low-temperature particulate removal costs were lost in overall cleanup systems. A 'grass roots' venturi or packed bed scrubber costs under $5/ACFM(4), since a great amount of gas/liquid contact is necessary as part of the sulfur/ammonia removal, the costs assigned to the particulate removal would be appreciably less.

High-temperature particulate removal, however, requires the use of high-strength, high-temperature alloys for cyclone and pressure vessel construction. Also, since most of the collection mechanisms for the smallest particles require periodic cleaning, two systems are required if continuous overall system operation is desired. As part of the Energy Conversion Alternatives System (ECAS) study carried out by the General Electric Company and Westinghouse Electric, high-temperature

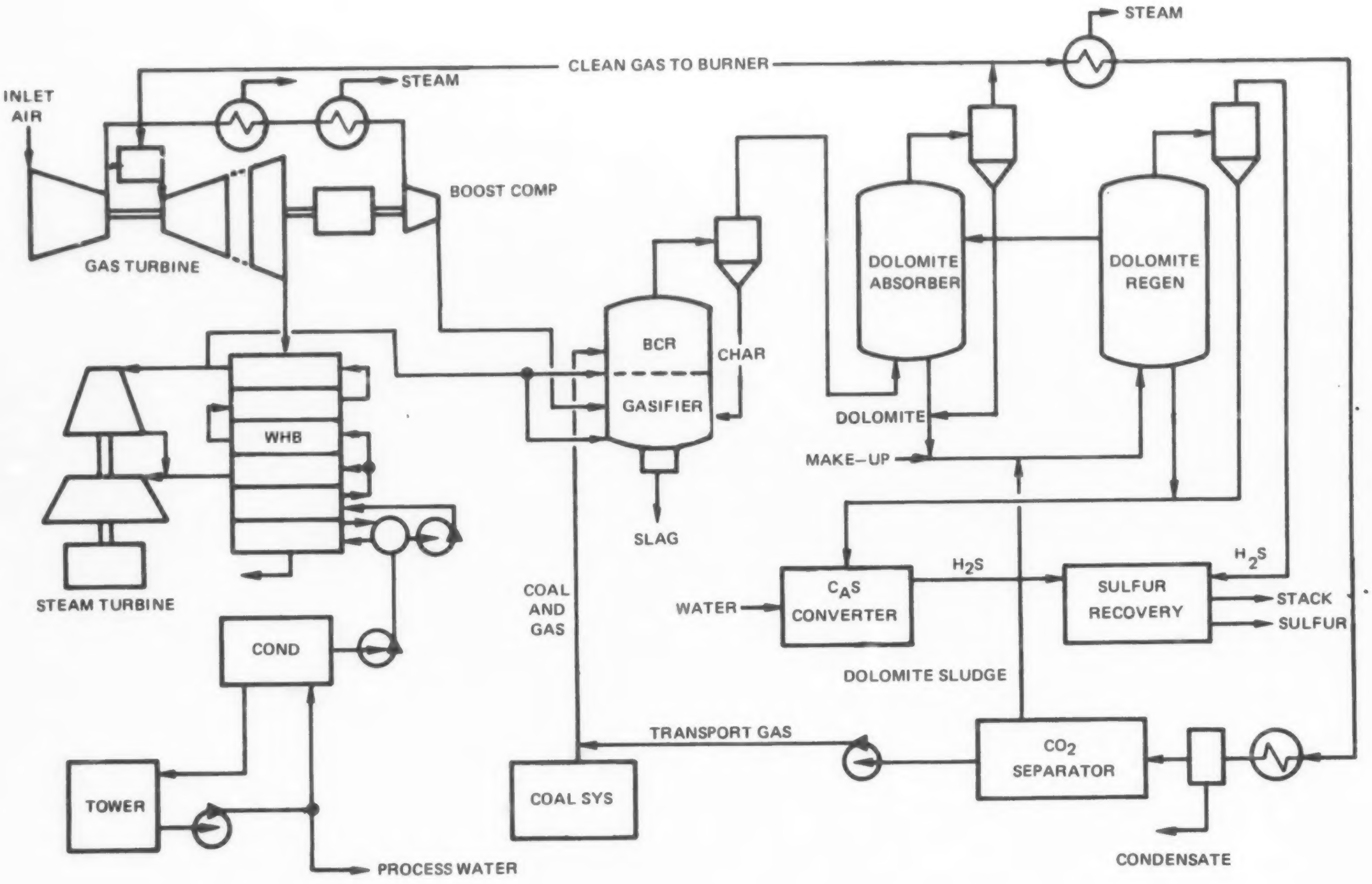

Figure 4: BCR/Conoco System

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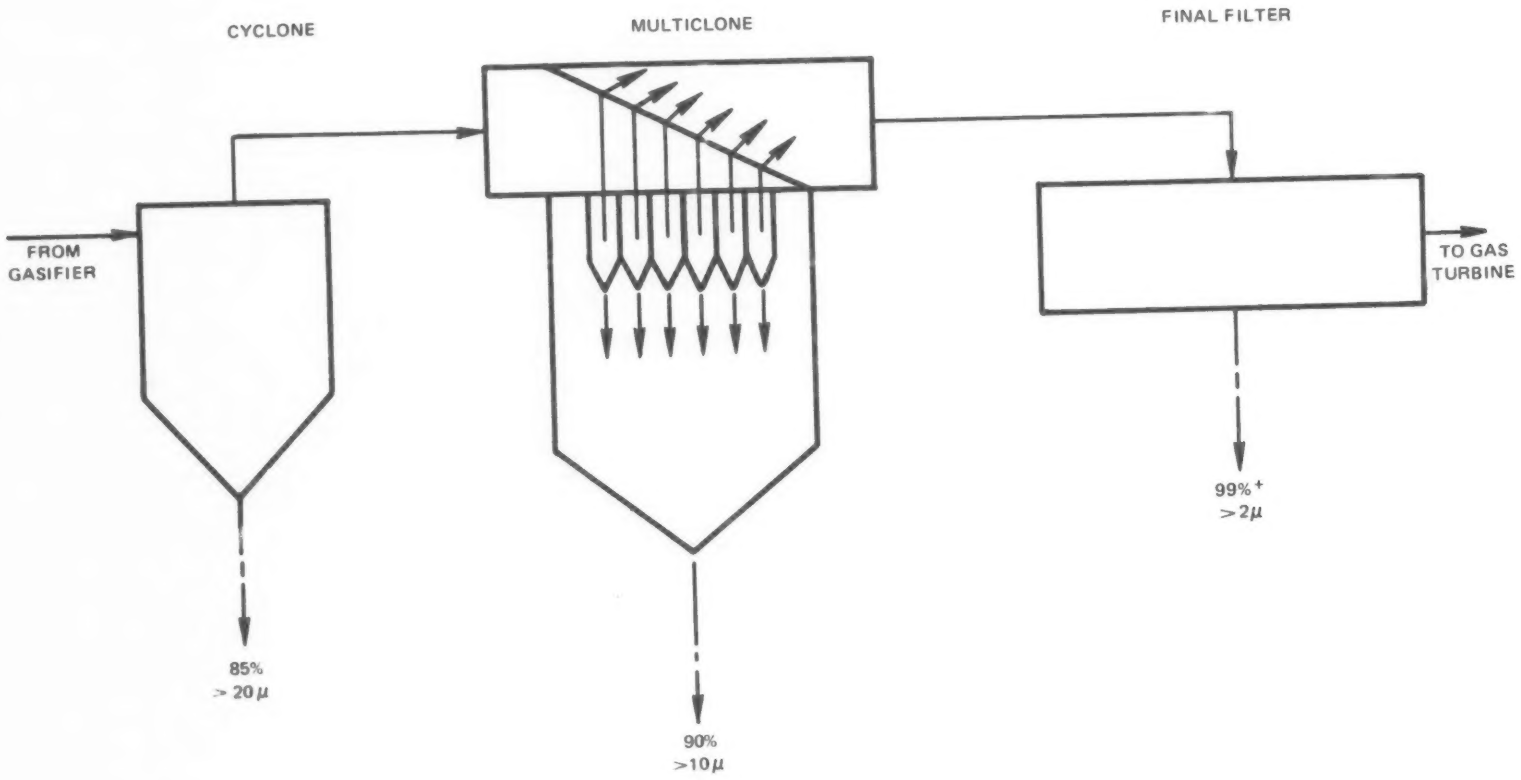

Figure 5: High–Temperature Particulate Removal System

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cleanup systems (cyclones, multiclones and granular bed filters) were investigated and costs projected.(5,6) The equipment along with the necessary installation, auxiliaries, piping, etc. was estimated to cost in the order of $200/ACFM, or, for a nominal 1000-MW plant over $20,000,000 some 5-7% of total plant cost. As part of an EPRI-sponsored study, Stone and Webster has indicated that a system of cyclones and metallic mesh filter would be in the same cost neighborhood.(7)

The effect of these costs are reflected in the cost of power from the high-temperature systems as shown in Table 4.

The Cost of NO_x Control

The estimated NO_x emission from the example integrated power plant using high-temperature sulfur and particulate control does not meet the EPA regulations (Table 4). The problem is two-fold: (1) NO_x due to thermal effects and (2) NO_x from fuel-bound nitrogen. The high fuel temperature (1600 F) precludes the use of premix burners; i.e., burners in which the fuel and combustion air are intimately mixed prior to entrance into the combustion chamber. Premixing could reduce thermal NO_x by 70% or more.(8) The fuel bound nitrogen is converted to NO_x during combustion on essentially a mol to mol basis. As can be seen in Table 4, over 5 lb/10^6 Btu is projected.

To reduce the NO_x and also to remove particulates and alkali metals, a third system has been included (see Tables 4 and 5). In this system a water wash is used subsequent to the high-temperature sulfur removal. The inclusion of the water wash allows the attainment of the EPA goals, but only at a performance and economic penalty.

RESEARCH AND DEVELOPMENT NEEDS

Without a better definition of both the particulate content of the fuel gas leaving the gasifier and the ability of the turbine to withstand particulate injestion, it is difficult to identify in any more depth the requirement for particulate cleanup. Thus, research in these areas is of first priority.

In addition, work on high temperature cleanup systems currently in the pilot stage should be reviewed with the better problem definition in mind, and those devices which show continued promise should be scaled to larger sizes.

Although the fuel bound nitrogen problem is associated with only a few of the gasifier types, it would be desirable to develop methods of removing nitrogen compounds at high temperatures.

CONCLUDING REMARKS

The potential advantages of high-temperature sulfur and particular cleanup; e.g., an approximately 10% lower electricity cost than a comparable power system with low-temperature cleanup, make such processes of great interest. However, problems associated with meeting the NO_x emission standards due to the fuel-bound nitrogen in the off gas from some gasifiers, may limit the application of high-temperature cleanup systems.

Much of the work on particulate removal devices which would operate at the 1600 F level is directly applicable to pressurized-fluid bed systems. Thus, there is a dual incentive to continue efforts in this area.

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